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An Evaluation of
Future Coal Technology

by
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# AN EVALUATION OF FUTURE COAL TECHNOLOGY

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## 1.0 INTRODUCTION

This research study was designed to be part of an input-output regional energy demand study of the Ohio River basin to the year 2000. In turn, the demand study is a supplementary part of the U.S. Environmental Protection Agency funded impact assessment of energy conversion facilities in the Ohio River Basin Energy Study (ORBES) region. The purpose of the present study is the establishment of the technical input-output coefficients for existing and probable future coal utilization technologies. These will be integrated into an I/O format to which subsequent alterations may be readily made. The present study entails a three-phase operation.

The first phase is the determination of those processes which can be commercially available by the year 2000; the basic constraint is the time horizon. For example, Figure 1.1 (1,p.935) demonstrates the normal procedure involved in bringing a coal gasification technique to commercial application. Some approximate relative costs are also shown. The time spans for plant testing have been shown, based on past performance, to be rather optimistic. They appear to represent a minimum time requirement. Therefore, the initial list of coal conversion processes includes only those that are already being tested at least on a pilot plant scale.



> "Meeting the nation's energy needs in 2000 requires commercial demonstration by 1985 of a spectrum of first and second-generation technologies for the production of high and low Btu gas, distillate, residuals, solvent-refined coal, and methanol 'so that informed investment and environmental decisions for achieving the 2000 goal can be made'." (2,p.76).

This is illustrated in Table 1.1 (2.p.76)

The initial list of possible processes is rather large. However, all must be considered feasible by the year 2000. These processes are shown in Tables 1.2, 1.3, and 1.4. Much of the information for these tables was obtained from Sargent and Lundy (3). This was updated where needed.

The second phase is the detailed cost analysis of the various techniques. This involves not only projections of the original capital costs, but an analysis of the relative operating costs of a commercial-sized plant. The magnitude of the various problems, and their consequences, are illustrated for the various processes. The detailed analysis included in the first phase is augmented in the second. Finally, the most realistically available techniques are chosen based on the economics involved.

Tables 1.2, 1.3 and 1.4 deal only with coal gasification and liquefaction. Obviously there are other coal utilization

technologies: MHD, fuel cells, combined cycle generation, and direct pyrolysis. However, much of the work in these areas is dependent on prior improvements in coal gasification; improvements in conjunction, or competition, with coal liquefaction.

Tables 1.2, 1.3 and 1.4 demonstrate the fundamental breakdown of the processes considered; low and high Btu gasification, and coal liquefaction. Separating low and high Btu gasification essentially differentiates between the primary characteristics in gasification processes: type of reactor, coal handling, and scale of use. Coal liquefaction techniques are maintained under a single heading because the major difference in these processes is between indirect and direct hydrogenation with Fisher-Tropsch the only major process using indirect hydrogenation.

A synopsis of the processes listed in Tables 1.2-1.4 is given, graphically when available, as well as their potential products. But the primary purpose of the analysis is to set forth those considerations which influence the feasibility of their commercial development. In conjunction with this, two different ratings are given in the tables. The first is the Status Rating (SR). This is a relative listing of the availability of the processes according to their 1978 status. The general breakdown of the rating system is:



1. Commercially available and used (or under construction).
2. Commercially available but not used.
3. Demonstration plant testing
4. Demonstration plant construction.
5. Pilot plant testing

The second, more subjective, rating is the Feasibility Rating (FR), which compares the feasibility of the processes in relation to the primary advantages or the shortcomings involved in conversion. The guidelines for this analysis are:

1. Refinement of current process
2. Problems most likely to be resolved
3. Testing limitations of process
4. Process used regardless of shortcomings
5. Problems inherent to process and not used at present

Not all of the specific processes listed are discussed here because some of them are merely extensions or combinations of other methods. Furthermore, those processes for which there was

insufficient data for comment were simply listed. In many cases, where data are unavailable, the omission is not due to proprietary disclosure but, rather, to the fact that the process is, at best, in a very formative stage.

It is rather obvious that the above discussion is preoccupied with coal conversion technologies. This will also later be evidenced by the separate comment section for the coal conversion chapter. However, this is because many of the technological and economical considerations involved in coal conversion reflect directly on other coal utilization technologies. For example, the aforementioned MHD process with coal is dependent on an efficient gasification process. This dependency is especially true of the high-sulfur coal in this geographic region of study. Consider direct combustion of coal in a fluidized medium - this could be an efficient way to handle the caking coals of this region. In addition, economic ways of handling the sulfur products could make combustion environmentally desirable. But hot-gas cleanup of coal gasification products would make a combined cycle system attractive. There is then the relative costs of transporting this power to and/or throughout the region. These alternatives are all considered in subsequent chapters, but the future of coal utilization - specifically high-sulfur, caking coal in the ORBES region - is considered with coal conversion technology as a baseline.



TABLE 1.1

Plants Required for Gas-from-Coal Industry

| By 1985 | Unit Plant Output | No. Plants | Unit Cost (1977 dollars) | Capital Investment $ billion (1977 dollars) |
|---|---|---|---|---|
| Pioneer and demonstration plants | | | | |
| High-BTU pipeline gas | | | | |
| Lurgi technology | 250 MMcfd | 3 | 1.4 billion | 4.2 |
| Advanced second-generation technology | 80 MMcfd | 3 | 400 million | 1.2 |
| Low BTU gas | 50 billion BTU/day | 3 | 250 million | 0 75 |
| Total | | | | 6.15 |
| By 2000 | | | | |
| High-BTU gas plants | 250 MMcfd | 60 | 1 1 billion | 66 |
| Low-BTU gas plants | 150 billion BTU/day | 30 | 600 million | 18 |
| Total | | | | 84 |

Source: Gas Research Institute (2.p.76)

TABLE 1.2

Low-Btu Coal Gasification Status

| PROCESS NAME | OFFERED BY | PROCESS SITE | PROCESS REMARKS | STATUS REMARKS | COAL TYPE | PROJECT SIZE | BTU | EFF. | $/MMBTU | $O_2$ NEEDED | REACTOR PRESSURE | SR | FR |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| KOPPERS-TOTZEK | KOPPERS | --- | $O_2$ Gasification Suspension Bed | Commercially Available | --- | --- | 302 | 77 | --- | Yes | Atmos. | 2 | 5 |
| LURGI | LURGI (Am. Nat. Gas) | N. Dakota (const.) | Fixed Bed Nonslagging. | Commercially Available | Sub-Bit., Lignite | 250 MMscfd | 302 | 78 | --- | Yes | 300-500 | 1 | 4 |
| U-GAS | IGT | Chicago | Fluidized Bed Ash-Agglomerating. | Pilot (testing) | Caking | 840 TPD | 136 | 71 | --- | No | --- | 5 | 2 |
| WELLMAN-GALUSHA | McDOWELL-WELLMAN | York, Pa. (const.) | Fixed Bed Gasifier | Commercially Available | --- | 150 TPD | 160 | 80 | --- | --- | Atmos. | 2 | 4 |
| WINKLER | DAVY POWER GAS | --- | Atmospheric Fluidized Bed | Commercially Available (demo. proposed) | All U.S | 8 TPH | Low/Med | 91 | --- | Either | Atmos. | 4 | 3 |
| ADVANCED GASIFICATION SYSTEM FOR ELECTRICAL POWER GENERATION | WECO | Waltz Mill, Penn. | Pressurized Fluidized Bed Gasifier | Scale-up in Design | Caking | 15 TPD | 130 | --- | --- | No | 15 | 5 | 3 |



TABLE 1.3

High-Btu Coal Gasification Status

| PROCESS NAME | OFFERED BY | PROCESS SITE | PROCESS REMARKS | STATUS REMARKS | COAL TYPE | PROJECT SIZE | BTU | EFF. | $/MMBTU | $O_2$ NEEDED | REACTOR PRESSURE | SR | FR |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| BI-GAS | Bituminous Coal Research | Homer City, Pa. | Coal Reacted with $H_2$-Rich Gas; Two Stage Gasifier; Carbonization; Entrained Bed Reactor | Pilot (Testing); Problems with Fines in Coal Feed | All U.S. | 120 TPD | 943 | 69 | 3.50 | Yes | 1000-1500 | 5 | 2 |
| $CO_2$ ACCEPTOR | Con. Coal Dev. Co. -Conoco | Rapid City, So. Dakota | Coal Gasified with Steam in Presence of Dolomite; Devolatilization; Fluidized Bed Reactor | Pilot Complete in 1972. Operational Problems | Sub-Bit. Lignite | 40 TPD | 953 | 63 | 3.50 | No | 150-300 | 5 | 3 |
| COGAS | ICCG | Cutler, Ill. | Multi-Stage Pyrolysis. Carbonization. Fluidized-Bed Reactor. Off-Shoot of COED. | Demo. Authorized | All U.S. | 2000 TPD | 950+ | 66 | --- | No | 50 | 4 | 2 |
| HYGAS | IGT | Chicago | Hydrogasification. Fluidized | Operating Pilot | All U.S. | 75 TPD | 940+ 947* 941** | 71* 57** | 2.75 | Yes* No*** | 1000-1500 | 5 | 2 |
| SLAGGING LURGI | Con. Coal Develp. Co. | Noble, Ohio; Westfield, Scotland | Modified Lurgi | Demo. Authorized; Pilot | All U.S. | 3787 TPD | 950+ | 63 | --- | --- | --- | 4 | 4 |
| SYNTHANE | Lummus Co. | Bruceton, Pa. | Two Zone Fluidized Bed Reactor. Devolatilization and Hydrogasification. | Pilot (testing) | All U.S. | 72 TPD | 927 | 65-70 | 4.50 | Yes | 1000 | 5 | 2 |

*Steam-$O_2$
**Steam-Iron
***Electrothermal and Steam-Iron

TABLE 1.4
Coal Liquefaction Status

| PROCESS NAME | DEVELOPED BY | PROCESS SITE | PROCESS REMARKS | STATUS REMARKS | PROCESS TYPE | PROJECT SIZE | EFF. | PRICE OIL,GAS | SH | FR |
|---|---|---|---|---|---|---|---|---|---|---|
| BERGIUS | --- | --- | Commercially Available | No longer used | Hydrogenation | --- | 55 | --- | 1 | 5 |
| COED | FMC | Princeton N.J. | Multistage Pyrolysis. Fore-Runner of COGAS. | Process Demonstrated on Pilot Plant. Pilot Work Terminated. Demonstration Unit Under Design. | Pyrolysis | 36 TPD | --- | --- | 4 | 2 |
| CLEAN COKE | U.S. Steel | Monroeville, PA. | Fluidized Bed. Combination Carbonization and Hydrogenation. | PDU Complete. Four TPD Pilot Plant Decision Pending | Pyrolysis and Hydrogeneration | 1/2 TPD | --- | $13/BBL | 5 | 2 |
| DONOR SOLVENT | Exxon | Baytown, TX. | Solvent Extraction | Pilot (under const.). One TPD PDV is Operational at Baytown, TX. | Solvent Extraction | 250 TPD | 65-75 | $20-25/BBL | 5 | 3 |
| FISCHER-TROPSCH | --- | U.S. Africa | Synthesis Gas Process | Commercially Available | Synthesis | --- | 55-70 | $13.05/BBL $2.25/MCF | 1 | 5 |
| H-COAL | Hydrocarbon Research | Cattlesburg, KY. | Similar to SRC. Ebullated Bed Reactor. Modification of H-OIL | Process Demonstrated on 3TPD PDU Demo. (const.) | Catalytic Hydrogeneration | 600 TPD | 60-65 | $14-18/BBL | 4 | 2 |
| SRC | PAMCO/GULF Energy and Mineral Co. | Tacoma, Wash.; Morgantown, West Va.; Owensboro, Ky. | Coal Slurry | Pilot Operating since 1974. Liquid (KY) and Solid (W. Va.) SRC DEMO. Plants Proposed for 1983. | Solvent Extraction | 50 TPD; 6000 TPD; 6000 TPD | 60-70 | $15/BBL $2.30/MCF | 4 | 2 |



FIGURE 1.1

| ITEM | SMALL PROCESS TEST UNIT | LARGE PILOT PLANT | DEMONSTRATION PLANT | FIRST GENERATION COMMERICAL PLANT |
|---|---|---|---|---|
| PLANT OPERATIONS: DESIGN, CONSTRUCTION & START-UP, OPERATION (TYPICAL TIME SCHEDULE - YEARS 0–16) | | | | |
| GASIFICATION CAPACITY (MMCF/DAY) | VARIABLE | 1.0–2.5 | 50–100 | 250+ |
| COAL FEED: TONS PER DAY / TONS PER YEAR | 1–3 / 50–150 | 50–125 / 15,000–50,000 | 2,500–5,000 / 1,000,000–2,500,000 | 12,500+ / 5,000,000+ |
| PLANT CAPITAL INVESTMENT ($ MILLION) | $0.5–1.5 | ($5.0–15.0) $10.0–30.0 | ($100–150) $150–250 | $150–250 (OR EXPAND DEMO. PLANT AT A COST OF $100–150) |
| OPERATING COSTS ($ MILLION/YEAR) | $0.5 | $2.5–4.5 | $20–35 (WITHOUT CREDIT FOR GAS SALES) | $50+ |
| APPROXIMATE TOTAL PROGRAM COSTS ($ MILLION) | $2–5 | $15–35 | $150–300+ | OFFSETTING REVENUES & VARIABLE |

Timetable for R & D Work to Culminate in Commercial Applicability of Coal Gasification Facility

Source: (1,p.935)

## REFERENCES

1. "Coal Technology." Modern Energy Technology, Vol. 2, Research and Education Association 1975.

2. Linden, H.R., "All-out push urged for U.S. gas supply " The Oil and Gas Journal November 21, 1977.



# 2.0 COAL CONVERSION

## 2.1 Coal Gasification

In general, coal gasification processes involve the following steps:

- Transport
- Preparation
- Preheat
- Gasification
- Tar Removal
- Shift Conversion
- Purification
- Methanation
- Transport and Storage

The principal differences among coal gasification processes are the manner in which coal is admitted to the gasifier, the type of reactor bed, and the gasification heat source. These differences may be dictated by the type of coal available. A generalized flow diagram with accompanying reactions is shown in Figure 2.1 (5.p.45) for the gasification process. If air is used in the hydrogasification step, a low-Btu gas is the result; if

oxygen is used, an intermediate-Btu gas is the product. If a high-Btu gas is desired, shift conversion and methanation steps are required. One of the desirable features of coal gasification is the ability to purify the gas with proven technology by removing the $H_2S$. The main disadvantage of this feature, however, is that currently the gas must first be cooled, which lowers the overall efficiency of the process.

Figure 2.2 (6,p.27) locates major water and coal resource regions, while Figure 2.3 (6,p.33) indicates potential areas for coal gasification. The ORBES region has abundant high-sulfur coal reserves for both deep and strip mines. Overall there is a large developed mining industry and a considerable market for a coal gasification product. However, the economic factors are at least as important as raw-material existence. Large blocks of reserves, such as bituminous coal in Illinois, would help keep costs down for such a capital-intensive process [39].

### 2.1.1 Low and Medium-Btu Coal Gasification

> "There are three main reasons for developing low-Btu gas technology: to develop a gasifier that can economically handle large volumes of gas throughput; to achieve high conversion of coal to gas without any liquid by-products; and to develop a process that will gasify coal with a minimum of damage to the environment." (7,p.51)



Low and medium-Btu gasification processes are seen primarily as an inexpensive and clean utility boiler fuel. The processes have some inherent advantages [49]:

1. They can produce gas using state-of-the-art technology.
2. They are less complicated and hence, less expensive than high-Btu processes.
3. They are amenable to modular expansion and ease in sizing.
4. The gas cleanup can be designed directly into the system and meet present standards.

Electric utilities, large-scale plants, and industrial parks, are feasible applications for a low-Btu gasification product. For example, there are no major technological constraints to the use of a medium-Btu gas in a secondary steel plant or in using a low-Btu gas in a refinery [30]. A study by Gilbert Associates (8,p.4) has shown that low-Btu gasification is the most economical coal conversion process for industries requiring up to 50 million Btu/day. For uses requiring more than this, a fixed bed pressurized system producing medium-Btu gas becomes lower in cost. At 150 billion Btu/day, an atmospheric entrained bed gasifier producing medium-Btu gas appears to be the most economical. For an output of 100 billion Btu/day, they found that a low-Btu fixed bed gasifier has the lowest initial cost, with the operating costs for fixed-bed, fluidized-bed, and

entrained-bed systems being about the same. Four of the feasible techniques listed in Table 1.2 are commercially available. Three of them are of the fixed bed variety. This is indicative of the initial emphasis placed on utilizing second generation techniques for producing a "pipeline-quality" gas.

2.1.1.1 Lurgi (9,10,11,28,35,48)

This technique is a fixed bed, pressurized, gasifier developed in Germany some 40 years ago. It is the archetypical first generation technique. Figure 2.4 (11,p.80) illustrates the coal path. High-pressure operation enhances methanation.

Scaling up is a major obstacle for this process. Additionally, the fixed bed process has trouble with eastern caking coals and coals containing fines. The temperature must be strictly controlled so that the ash fusion temperature is not reached. This limits process efficiency and explains why current plans for construction are all done outside the ORBES region where non-caking coals are abundant (28,46). However, world-wide, many different coals have been used with this process. In addition to the above, leading to high initial investment, operating costs are increased by the required use of oxygen. However, air could be used to produce a raw gas with the



composition shown in Table 2.1 (11,p.80).

2.1.1.2 WELLMAN-GALUSHA (11,12,48)

This is another fixed bed reactor process that operates at atmospheric pressure with an agitator to help maintain a uniform bed. Its raw gas composition is shown in Table 2.2 (11,p.87).

All types of coals can be used if the agitator is applied but, again, this system must be operated below the ash fusion temperature, reducing efficiency. Its low operating pressure limits its possible utilization. Although this gasifier has been operated commercially for many years, its limited reactor size deters future applications. Tests have been run on a pressurized Wellman-Galusha gasifier, but efficiency is still limited by fusion temperature.

2.1.1.3 KOPPERS-TOTZEK (6,9,10,11,48)

This first generation, proven process, was developed in 1948. Although no plants exist in this country, this suspension-bed process seems to have some advantages over Lurgi.

Coal size is not a limiting factor, and caking coals can be handled without pretreatment. Figure 2.5 (11,p.81) illustrates the entire process. Figure 2.6 illustrates a simplified design of the reactor. The slagging system increases throughput over a fixed-bed system. In addition slagging temperatures increase steam conversion and pressurization enhances methanation.

Fly-ash removal has been a problem with this technique. The major drawback is that it requires almost 30 percent more oxygen than the Lurgi process for an equivalent amount of $CO + H_2$. Air cannot be used because temperatures would not be high enough for a slagging operation. Again scale-up could be a problem. Typical raw gas compositions are shown in Table 2.3 (11,p.81).

2.1.1.4 WINKLER (10,11,17,48)

The Winkler process is the only commercially-proven fluidized bed process. It was first proposed in 1922. Its advantage is that this process can be used on caking and high ash coal. Furthermore, no tars or oils are produced. The process is relatively easy to operate. However, it is an atmospheric process, and entrained coal must be recycled for good efficiency. As in all fluidizing processes, coal reactivity is a strong determinant in process efficiency. Partial oxidation may be



required for caking coals. A flow diagram of the process, and its raw gas composition, are given in Figure 2.7 and Table 2.4, respectively (11,p.86).

2.1.1.5 U-Gas (3,7,11)

This is one of the few second generation fluidized bed techniques considered feasible for low-Btu gasification by the year 2000. The process flow is shown in Figure 2.8 (7,p.51).

This technique has the desirable low-Btu process features of not producing tars and oils, having a simple design, and the ability to handle all types of coal. However, pretreatment is required for caking coals. A drawback of many fluidized systems is their inability to handle low carbon-content ash. This is essential in order to maintain non-slagging operation. Thus, pilot work is being done whereby the ash-agglomerating technique (AAG) is incorporated into the gasifying region of the process. This process is shown in Figure 2.9 (7,p.53), with Table 2.5 (7,p.53) listing AAG operating data. To date all tests have been done only at atmospheric pressure. Modifications have been made for accepting coal rather than char. The process would then be as shown in Figure 2.10 (7, p.52).

2.1.1.6 Advanced Gasification System (3,11,29,31)

A second generation fluidized bed process that may be desirable is that developed by the Westinghouse Electric Corporation. It seems to have the distinct advantage of being able to handle eastern caking coals. The process is shown in Figure 2.11 (11,p.87) with typical raw gas composition in Table 2.6 (11,p.88). Coal is blown through an open-ended concentric "draft tube" inside the gasifier vessel. The coal suspension plus the use of this concentric tube presumably keeps the coal from caking. In addition, at these high temperatures the ash agglomerates until they fall from the suspension. Caking deterence and ash removal eradicates two major drawbacks of using eastern coals. Dolomite is used in the suspension for $H_2S$ removal. Effective scale-up (in design now) would make this process very desirable and applicable in many situations.

2.1.2 High-Btu Gasification

High-Btu gasification produces a clean synthesized form of pipeline quality (950+ Btu/scf) "natural" gas. This gas could be used for industrial and residential consumption. The currently high operating costs of these rather complex processes make the gas relatively expensive. Coal feed and $O_2$ requirements are



common problems to most of the processes. These processes are almost exclusively second-generation fluidized and entrained-bed gasifiers, indicating that the original emphasis in gasification technology was placed on high-Btu gas production.

2.1.2.1 BIGAS (1,3,4,5,9,11,13,31,48)

This technique uses a high-pressure, two-stage, gasifier as shown in Figure 2.12 (13,p.56). The reactor operates at rather extreme temperatures (2700°F), which may contribute to materials problems. The high pressures involved enhance the methanation process. The slagging removal system should be noted especially. No by-products are produced that require additional processing. Again, no tars or oils are produced.

This entrained bed system is said to have the advantage of being able to handle all types of coal. The overall process is shown in Figure 2.13 (11,p.75). However, a large quantity of fines are produced. This has prompted the performance of alternate coal preparation tests. These alternate paths are shown in Figure 2.14 (13, p.55).

Many of the problems still associated with this process, such as the coal preparation change mentioned above, are in the

pilot plant testing stage. Other tests involve water-cooling flows, materials testing, and ignition design. Operation of the process could be complex. The plant is essentially in the start-up phase; various pressurized tests are still required. This process seems to be best suited for eastern bituminous coals, although high ash fusion coals could be a problem for the slagging operation. Unfortunately, it has a large $O_2$ requirement, even though this requirement is reduced by preheating the fuel. Pressurized operation with air has not been demonstrated.

2.1.2.2 $CO_2$ Acceptor (1,3,4,5,9,11,31,34)

This process, shown in Figure 2.15 (11,p.76), is desirable because it does not require an external source of oxygen or hydrogen, reducing both initial and process costs. The $H_2/CO$ ratio of purified gas is approximately 3:2, eliminating the need for shift conversion.

Materials handling and dolomite regeneration have caused some problems. In fact, operation of the process has proven quite delicate. However, the major drawback to this process is that it only operates effectively with the highly reactive lignite and sub-bituminous coals.



2.1.2.3 COGAS (9)

This conversion process is an outgrowth of the COED liquefaction process discussed below. It is being converted from char gasification to coal. Materials handling is one of the major operational problems involved.

2.1.2.4 HYGAS (1,3,4,5,9,11,14,15,16,31,33,51)

Rather extensive testing has been done on this process to determine its operating characteristics as a slurry-to-fluidized bed multi-stage gasifier technique with eventual scale-up to commercialization. All types of coal can be used, although it naturally operates best with highly reactive lignite. A 92 percent coal conversion has been attained with Illinois #6 coal. Mild-air oxidation pretreatment is required for certain types of coals in order to destroy the coal's agglomerating tendencies. The diagram in Figure 2.16 (15,p.75) illustrates the process.

The expensive aspect of this process is its need for a hydrogen-producing process (electrothermal, $O_2$-steam, steam-iron). The process shown in Figure 1.17, being studied by IGT, utilizes the $O_2$-steam generation system. There is less need for

methanation in this process compared to others. Table 2.7 (15,p.76) shows some of the operating characteristics and the relative raw gas yields.

This process is a representative example of those second generation processes where there might be a decision required between availability and cost. Tables 2.8 (33,p.29) and 2.9 (33,p.10) illustrate some of the costs associated with this process for a commercial 250 MMscfd plant using Pittsburgh seam coal.

2.1.2.5 Slagging Lurgi (3,10,52)

This process, as its name suggests, is an extension of the proven Lurgi technique. The major difference is the higher operating temperature (fusion temperature is restrictive in the original Lurgi process) for reducing present ash to slag for removal. This higher temperature improves the overall thermal efficiency of the process and, hopefully, will enable the process to handle high-agglomerating eastern coals. This has been demonstrated on a pilot plant scale (52). However, many of the problems inherent to the Lurgi process, such as scale-up, are still present in this, essentially first generation, process. Process comparisons with Lurgi are made in Table 2.10 (10,p.74).



2.1.2.6 SYNTHANE (1.3,4,5,6,9,11,18,31.32)

The Synthane process is a fluidized-bed gasification scheme being developed by the Bureau of Mines. If operated correctly, its inherent simplicity (no major recycle lines) could make it a favorable future technique. Figure 2.17 (18,p.81) is a detailed diagram of the procedure. External coal pretreatment is eliminated, and all types of coal should be applicable. Considerable methane is produced in the gasifier, which helps reduce methanation requirements. Table 2.11 (11,p.84) lists some typical raw gas compositions.

During testing there has been a problem with fines and clinkers. Oxygen requirements are still present, but are lower in relation to comparable processes. Materials testing is still required due to sulfur related corrosion problems. Many of these current problems should be resolved during pilot testing. Synthane would then be one of the more promising second generation techniques, with relatively inexpensive operating costs. Tables 2.12 (32,p.11) and 2.13 (32,p.24) list some representative capital and annual operating costs, respectively, for a hypothetical 250 MMscfd plant using Pittsburgh seam coal.

## 2.2 Coal Liquefaction

As stated above, the initial emphasis of coal conversion technology was aimed at developing a high-Btu gasification process, with low-Btu gasification applied to more localized operation. Indirectly, however, coal liquefaction has been drawing more and more support:

> "But in the opinion of knowledgeable researchers in the government, the optimum path to be pursued is not one of coal gasification, nor even coal liquefaction, although for many reasons liquefaction is more attractive than gasification, but that of a multiproduct production .... In a plant of that type, the entire potential chemical energy of the coal would be employed usefully down to the last bit." (9,p.275)

Coal liquefaction is a relatively new technology, although it was first developed in Germany in the 1920's and 1930's. One of the original techniques, Bergius, is no longer in operation anywhere in the world. The other, Fischer-Tropsch, has survived and is now used commercially (although not in the U.S.), most notably at the SASOL plant in South Africa. Some of the potential products of liquefaction by this process are exemplified in Table 2.14 (3,p.16). As can be seen, coal



liquefaction products could affect many markets, such as that of the automotive industry [37,38]. In fact, coal liquefaction's greatest advantage might be as a supplier of a synthetic crude oil:

> "Production of synthetic crude allows it simply to be added to the natural crudes still available to refineries. With relatively minor modification to the refineries, final products essentially identical to present fuels result. This approach has the practical advantage of serving both the needs of oil companies wishing to maintain the usefulness of present investments and of insulating the consumer from change." [47,p.xii]

Coal liquefaction is accomplished by the addition of hydrogen (directly or indirectly) which results in the breakdown of the long and complex hydrocarbon molecular chains. Usually hydrogen is produced by a gasification or solvent process. Desulfurization usually occurs during a gasification step. There are four principal processes used to accomplish liquefaction (50.p.VII-5):

1. Hydroliquefaction (direct catalytic hydrogenation)

2. Solvent extraction (noncatalytic liquid phase-dissolution)

3. Pyrolysis

4. Liquid hydrocarbon catalytic synthesis (Fischer-Tropsch-indirect hydrogenation)

Hydroliquefaction and solvent extraction appear to be the most promising techniques for second generation processes. As the majority of the liquefaction processes use a form of hydrogenation, the primary differences among the processes occurs due to the amount of hydrogen required, catalyst requirements, and operational difficulties. These are all dependent on the type of multiproduct fuels demanded from, or available in, the coals used.

2.2.1 Fischer-Tropsch (3,10,19,20,50)

As in the Lurgi process, the current advantage of this process is its commercial availability. A typical line diagram for the process is shown in Figure 2.18 (11,p.77), with possible product composition given in Table 2.15 (11,p.78).

This technique uses an indirect liquefaction process, whereby coal is gasified to carbon monoxide and hydrogen, which are recombined catalytically to produce a variety of liquid and gaseous fuels. Various gasification processes could be used with this process although, if used in the U.S., it would be preferable to use a technique which can handle caking coals (at present those used do not). Another deterrent may be the



abundance of by-products unsuitable for fuel, such as waxes, although cleaner fuels are obtained here than with direct hydrogenation.

If used now, the economics of this process would compare favorably with direct hydrogenation processes. In the long run, the operation of a direct hydrogenation process would be cheaper. Thus, some feel the SASOL plant should be used as a model while others argue that these "antiquated" gasifiers should be ignored and the U.S. should concentrate on making available the more modern processes.

2.2.2 COED (1.3.9.11,50)

This process, shown in Figure 2.19 (11,p.77) liquifies the coal by first passing it through a series of pyrolyzing reactors. This was the first successful pilot process.

The residual char was first thought to make a good boiler fuel, but the demand for pipeline quality gas has led to development of a gasification process utilizing this char, with extensions to the use of coal. This outgrowth is the COGAS

process mentioned above.

This type of plant has been operated with Illinois, Colorado, and Wyoming coals. Temperature control in the various gasifiers is one of the major operating considerations. An oxygen plant is required, which is typical of most liquefaction processes. Table 2.16 (11,p.77) lists some representative yield data using Illinois #6 coal.

2.2.3 SRC (1,3,8,11,16,19,23,24,36,40,41,50)

Solvent refined coal is one of the original solvent extraction processes. It was originally developed for producing sulfur free coal. However. this is not its exclusive product.

> "Only when product yields are placed in the perspective of everyday units of the using industry is it clear that the co-product streams are far from insignificant. The SRC process might more properly be called a multi-synthetic-fuels process." (21,p.89)

In fact, funding is currently being allocated to produce both liquid and solid SRC (22,p.7). It is hoped it can be ready for commercial application by 1987. The process line is shown in



Figure 2.20 (11,p.83).

In this process the coal is dissolved in a coal-based internally-generated solvent while adding hydrogen under pressure at a temperatures of 800°F. Hydrogen and other gases are released in the pressure letdown. If solid SRC is required, the slurry is cooled. Variations in hydrogen addition are required to produce more potential liquid fuels. It would seem that this slurry should be fed directly into a refining process; i.e., incorporate the process into a "coalfinery." This would appear to give a little more latitude in product composition. Some typical SRC process yields are shown in Tables 2.17 (21,p.69) and 2.18 (21,p.89).

An advantage of this process is the relative experience in using this technique. However, a drawback is the amount of hydrogen required. A potentially greater problem with the liquid fuel is meeting $SO_2$ emission requirements. Furthermore, this process does nothing to eliminate $NO_x$ emissions. Tests are being done to study pollutant levels. A solvent deashing technique is being tested to remove sulfur and ash. Nevertheless, the many possible products from the SRC process may still make it one of the most desirable. Some representative costs are shown in Tables 2.19 (24,p.11) and 2.20 (24,p.34) for a 50,000 bpd SRC plant using Illinois #6 coal.

### 2.2.4 SYNTHOIL (3,8,11,36,50)

This is a hydrodesulfurization process being developed to produce low ash, low sulfur fuel oil. The flow process is shown in Figure 2.21 (11,p.84).

As in many of these processes, coal is slurried with some of its own coal-based oil with the subsequent addition of more hydrogen at elevated pressure and temperature. It is then fed to a packed bed catalytic reactor. Typically, external hydrogen is required, but an additional catalyst (such as cobalt molybate on silica-activated alumina) is required in the reactor.

### 2.2.5 Exxon Donor-Solvent (3,8,20,25)

A rapidly advancing technique is the donor-solvent process, in which coal is hydrogenated in a slurry generated by its coal-based solvent (oil). In this technique, however, the solvent is catalytically hydrogenated in a separate reactor before it is mixed with the coal feed. Hydrogen is supplied from the excess available in the donor solvent and from an external source. This may prove to be a very efficient process.



### 2.2.6 H-Coal Process (3,8,11,20,26,27,36,50)

The H-Coal process is a direct hydrogenation technique utilizing catalytic activation. The flow system is shown in Figure 2.22 (11,p.79).

This process is representative of the more advanced liquefaction technologies. An ebulliating bed of hydrogen, catalyst, coal, liquid and gaseous products is produced in the reactor and maintained at approximately 850°F and 2700 psig. Flash distillation at one atmosphere is used to separate light and heavy distillates. Typical products from Illinois No. 6 coal are shown in Table 2.21 (11,p.79).

The major drawback to the process is the extensive hydrogen requirement and, as in other processes, the cost of the catalyst. Tables 2.22 (27,p.12) and 2.23 (27,p.23) list some hypothetical costs for a 50,000 bpd H-Coal plant using Illinois coal.

## 2.3 Comments

It can be seen that liquefaction processes lean toward development of new techniques. This is also true for

gasification techniques, even though first generation techniques are being used in specific situations. The need for improvement is seen in the relative, and approximate, operating costs given in Table 2.15 (3,p.18).

Concerning the rating systems used in Tables 1.2-1.4, a shift is seen in emphasis from 1978 to the year 2000. In fact the "status rating" (SR) given relates to the present, whereas the "feasibility rating" (FR) is a rather subjective projection to the year 2000. The status rating is rather clear-cut and self-explanatory. It illustrates the state of the art. However, this could also be considered the feasibility rating for the year 1978, as the main criteria at this point in time is availability. There will be a steady shift in this feasibility rating in the upcoming years. By 1985, as the second generation processes become proven on a larger scale (and operating problems are resolved), it will be apparent that their availability is in the foreseeable future, and with more desirable characteristics.

Nevertheless a direction must be established for the aforementioned technologies. In selecting a frontrunner for the year 2000, alternative techniques are not discounted, but rather a configuration is established which will have the most probable significant impact on our fuel economy relative to its development. In doing so one can determine what hardware, and construction and operating costs, are required for a given flow



process. It should be pointed out that certain commercial applications are not advocated over others, but rather considered, reported, choices are made concerning the best overall technological process for the future. Upon establishing which concepts will predominate, the configurational and operational characteristics established by the appropriate group are used for an in-depth I-O cost analysis.

Consider Table 1.2. All technologies given in the table are worthy of consideration at present. The shift in ratings is indicative of the first and second generation processes. By 1985 there will be a phase-out of first generation processes, as second generation techniques gain economically and technologically. Note that modified first generation techniques will be in operation [43], but future construction plans should be shifting to newer processes. Lurgi will still have scale-up and caking coal problems, as will Wellman-Galusha. Koppers-Totzek will still have prohibitive $O_2$ requirements, and will probably still have fly-ash removal problems. This leaves the U-Gas and Winkler processes for the year 2000. Note that U-Gas is given a slightly higher rating. This is largely due to the fact that its testing has been the most extensive on a larger scale, and because of its use of the ash-agglomerating technique. Winkler is already a commercially proven process, and will feasibly be cost-competitive if $O_2$ is not used in the process. These processes are primarily desired for utility boiler

purposes, but they could also prove economically useful for a large industrial plant or an industrial park. Note that the future of low-Btu gasification technology lies with atmospheric fluidized bed processes. The development of U-Gas as a pressurized fluidized bed reactor process, resulting in increased efficiency, will make it the most desirable low-Btu gasification technique. This reactor is similar in concept to that of Westinghouse's Advanced Gasification System which was not discussed at length in this section simply because its development is in conjunction with a combined cycle system to be discussed later. Tables 2.24 (46,p.5-3) and 2.25 (46,p.5-2) list comparative costs for some general conceptual designs of second generation low-Btu gasification schemes for boiler application.

The transition in Table 1.3 is not as clear-cut simply because these are predominantly second generation gasification processes. However, those with the highest rating are those that have undergone sufficient testing to support optimism that the major problems with the process will be resolved or minimized. Again there is an elimination of some processes by 1985. The $CO_2$ Acceptor cannot be included because it may only be able to use lignite or sub-bituminous coal, which does not offer much of an environmental advantage and is not readily available in the ORBES region. In addition, slagging Lurgi will still have many of the problems inherent in the original Lurgi process. The COGAS process also will not be a factor past 1985, as it has materials



handling problems and will probably be used only as a char gasifier with its parent COED liquefaction process. BIGAS is also left out, primarily because of its extensive $O_2$ requirements. Methanation processes could possibly be available by 1985, but future plans will revolve around the remaining second generation techniques. These remaining processes for the year 2000 are Hygas and Synthane. Hygas has lower methanation step requirements, while the Synthane process has lower $O_2$ needs. In addition, no coal pretreatment is required for Synthane. Corrosion and fines problems are the major obstacles to be cleared. Note, that the most feasible processes utilize a fluidized bed reactor. However, while these seem to be the most economically and technologically attractive high-Btu processes, they are by no means inexpensive. Demand will ultimately dictate just how extensively they are used. For this reason the pipeline quality product gas will probably be used only for commerical/residential applications. The cost of production would seem to make it undesirable as a process fuel for industrial or utility uses. In fact, competition from low-Btu gasification and liquefaction processes may make its market too restrictive.

In Table 1.4 (liquefaction) some transition can be seen in the relative desirability of first and second generation processes. By 1985 Fischer-Tropsch, an indirect hydrogenation technique, will no longer be considered economically or

technologically competitive, as was the case with the original Bergius technique. The COED process will probably be shelved also as the pyrolysis method does not seem to be as efficient as the relatively newer methods. Thus, three desirable and increasingly popular [44] processes are left. The frontrunner at this time would seem to be SRC II. It is the most developed of the new techniques, and requires an internally-produced solvent. In addition two processes, H-Coal and Exxon Donor Solvent, are rated just behind in feasibility. This is mainly due to less development. H-Coal requires a catalyst, and Exxon, the support for which is increasing rapidly [42], requires both a catalyst and solvent. SRC is a solvent extraction technique, whereas H-Coal and Exxon are catalytic hydrogenation processes. Because all are multiproduct processes, the utilization is manifold. Even though hydrogen production is required for all processes, the various products make them desirable. In fact, coal liquefaction may have the best future of all coal technologies. These processes would seem to have the capability of providing commercially available help by the year 2000.



TABLE 2.1

Raw Gas Compositions for LURGI

| Component | Mol % High-BTU | Mol % Low-BTU |
|---|---|---|
| CO | 9.2 | 13.3 |
| $CO_2$ | 14.7 | 13.3 |
| $H_2$ | 20.1 | 19.6 |
| $H_2O$ | 50.2 | 10.1 |
| $CH_4$ | 4.7 | 5.5 |
| $C_2H_6$ | 0.5 | |
| Other than $H_2S$ | 0.6 | 0.7 |
| $N_2$ | | 37.5 |
| Total | 100.0 | 100.0 |
| Higher heating value (dry basis), BTU/scf | 303 | 180 |

Source: (10.p.80)

TABLE 2.2
Raw Gas Compositions for WELLMAN-GALUSHA

| Component | Mol % High-BTU | Mol % Low-BTU |
|---|---|---|
| $CO$ | 29.6 | 26.0 |
| $CO_2$ | 12.3 | 3.0 |
| $H_2$ | 30.3 | 13.9 |
| $H_2O$ | 25.3 | 8.3 |
| $CH_4$ | 0.7 | 2.5 |
| $N_2$ | 1.1 | 45.6 |
| $O_2$ | 0.1 | |
| $H_2S/COS$ | 0.6 | 0.7 |
| Total | 100.0 | 100.0 |
| Higher heating value (dry basis), BTU/scf | 268 | 168 |

Source: (11,p.87)



TABLE 2.3
Raw Gas Compositions for KOPPERS-TOTZEK

| Component | Mol% |
|---|---|
| $CO$ | 50.4 |
| $CO_2$ | 5.6 |
| $H_2$ | 33.1 |
| $H_2O$ | 9.6 |
| $CH_4$ | 0.0 |
| $H_2S+COS$ | 0.3 |
| $N_2$ | 1.0 |
| Total | 100.00 |
| Higher heating value (dry basis), Btu/scf | 180 |

Source: (11,p.81)

TABLE 2.4

Raw Gas Composition for WINKLER

| Component | High-BTU (Mol%) | Low-BTU (Mol%) |
|---|---|---|
| $CO$ | 25.7 | 19.0 |
| $CO_2$ | 15.8 | 6.2 |
| $H_2$ | 32.2 | 11.7 |
| $H_2O$ | 23.1 | 11.5 |
| $CH_4$ | 2.4 | 0.5 |
| $H_2$ | 0.8 | 51.1 |
| $H_2S$ | 2,500 ppm | 1,300 ppm |
| $COS$ | 400 ppm | 200 ppm |
| Total | 100.0 | 100.0 |
| Higher heating value (dry basis), BTU/scf | 275 | 118 |

Source: (11.p.88)



TABLE 2.5

AAG Operating Data

| | Air-steam feed, no fines recycle | | | | $O_2$-enriched air-steam feed with fines recycle | $O_2$-steam feed, no fines recycle | | | $O_2$-steam feed with fines recycle | |
|---|---|---|---|---|---|---|---|---|---|---|
| Run No.: | 10 | 31 | 36 | 48 | 31 | 39 | 47 | 48 | 51 | 53 |
| Number of cyclones | | | | | | | | | | |
| Internal | 1 | 0 | 0 | 1 | 0 | 0 | 1 | 1 | 1 | 1 |
| External | 1 | 2 | 2 | 2 | 2 | 2 | 2 | 2 | 2 | 2 |
| Average bed temperature, °F. | 1,872 | 1,884 | 1,900 | 1,895 | 1,920 | 1,940 | 1,910 | 1,928 | 1,940 | 1,960 |
| Coke breeze feed rate, lb/hr | 387 | 469 | 567 | 345 | 628 | 443 | 439 | 365 | 492 | 528 |
| Average solids feed mesh size | 14 | 14 | 16 | 16 | 14 | 16 | 16 | 16 | 16 | 16 |
| Venturi throat diameter, in. | 3 | 3 | 2 | 2 | 3 | 2 | 2 | 2 | 2 | 2 |
| Venturi throat velocity, ft/sec. | 89 | 99 | 151 | 146 | 80 | 246 | 172 | 167 | 109 | 112 |
| Total bed superficial velocity, ft/sec | 2.2 | 2.01 | 2.05 | 2.1 | 2.23 | 1.83 | 2.16 | 2.07 | 2.0 | 2.0 |
| Product gas heating value BTU/scf | 94 | 87 | 92 | 75 | 122 | 208 | 205 | 191 | 202 | 218 |
| Carbon gasification efficiency, * % | .. | 96.5 | 93 | .. | 97.8 | 96.5 | .. | 94 | 94 | 96.1 |
| Fines elutriation rate,‡ lb/hr | .. | 265 | 238 | 66 | 127 | 166 | 99 | 62 | 51 | 38.5 |
| Agglomerate production rate, lb/hr | 39 | 43 | 34 | 31 | 50 | 20 | 26 | 42 | 64 | 57 |
| Analysis§ | | | | | | | | | | |
| Carbon, % | 46.7 | 21.0 | 35.0 | 43.3 | 16.4 | 21.2 | 38.1 | 30.5 | 30.1 | 19.7 |
| Ash, % | 53.9 | 79.0 | 65.0 | 56.7 | 83.6 | 78.8 | 61.9 | 69.5 | 69.9 | 80.3 |
| Average agglomerate mesh size | No Agglomerate | 7 | 12 | No Agglomerate | 1/4 | 6 | No Agglomerate | 14 | 14 | 8 |
| Product gas composition (dry)‖, % | | | | | | | | | | |
| $CO_2$ | 10.3 | 14.3 | 13.9 | 14.7 | 12.6 | 38.6 | 39.0 | 42.7 | 39.2 | 34.5 |
| $CO$ | 12.2 | 15.5 | 16.6 | 12.0 | 21.8 | 24.6 | 24.9 | 21.3 | 23.3 | 29.8 |
| $H_2$ | 15.5 | 9.8 | 11.7 | 10.3 | 14.2 | 36.1 | 35.0 | 35.4 | 36.9 | 34.9 |
| $CH_4$ | 0.3 | 0.5 | 0 | 0.3 | 0.5 | 0.7 | 1.1 | 0.6 | 0.6 | 0.8 |
| $N_2$ | 61.7 | 59.9 | 57.8 | 62.7 | 50.9 | 0 | 0 | 0 | 0 | 0 |

†Calculated higher heating value based on gas analysis. * Based on carbon input compared to carbon lost in agglomerates. ‡Based on elutriated fines collected in first- and second-stage external cyclones, whether fines recycle was used or not. §Agglomerate carbon analysis includes any entrained fines collected with agglomerates. ‖Nitrogen purge-free basis.

Source: (7,p.53)

TABLE 2.6

Raw Gas Composition for "Advanced Gasification" Process

| Component | Mol % |
|---|---|
| CO | 17.7 |
| $CO_2$ | 8.6 |
| $H_2$ | 13.3 |
| $H_2O$ | 7.9 |
| $CH_4$ | 2.5 |
| $N_2$ | 50.0 |
| $H_2S$ | small quantity |
| Total | 100.0 |

Higher heating value (dry basis), 136 BTU/scf

Source: (11,p 88)



TABLE 2.7

HYGAS Pilot Plant Operating Characteristics

| | Period 1* | Period 2+ |
|---|---|---|
| Reactor operating conditions: Coal or char feed (dry), lb/hr | 5,912 | 5,478 |
| Process steam feed, lb/hr | 6,117 | 6,894 |
| Process oxygen feed, lb/hr | 824 | 960 |
| Reactor pressure, psig | 1,010 | 893 |
| Steam-oxygen zone temperature, °F. | 1,640 | 1,556 |
| Operating results Raw gas composition (dry basis), mole % $H_2$ | 22.73 | 26.65 |
| $CO_2$ | 34.30 | 35.30 |
| $C_2H_6$ | 0.47 | 1.39 |
| $N_2$ | 6.68 | 4.44 |
| $H_2S$ | 0.70 | 0 |
| $CH_4$ | 26.48 | 23.31 |
| CO | 8.63 | 8.91 |
| Total | 100.00 | 100.00 |
| Methane yield, scf/lb coal | 2.8 | 4.1 |
| CARBON OXIDE YIELD, scf/lb coal | 4.5 | 7.7 |
| Coal gasified, % | 53.0 | 74.0 |
| Carbon gasified, % | 45.0 | 67.0 |

*Period 1--Test 54, period 7/10/76 (0000 hr) to 7/11/76 (0000 hr); feed pretreated char from Illinois No 6 bituminous coal.

TABLE 2.8

**Total Estimated Capital Requirements and Operating Personnel for HYGAS Plant**

| Unit | Cost | Percent | Estimated number of operators per shift |
|---|---|---|---|
| Coal preparation | $3,270,200 | 0.7 | 2 |
| Coal pretreatment | 16,577,000 | 3.3 | 2 |
| Char slurry preparation | 5,111,800 | 1.1 | 2 |
| Hydrogenation | 78,413,500 | 15.7 | 4 |
| Dust removal | 3,497,700 | .7 | 2 |
| Waste heat recovery No. 1 | 5,595,600 | 1.1 | 1 |
| Shift conversion | 5,036,400 | 1.0 | 1/2 |
| Waste heat recovery No. 2 | 16,195,500 | 3.2 | 2 |
| B-T-X recovery | 6,083,100 | 1 2 | 2 |
| Purification | 61,510,000 | 12.3 | 5 |
| Methanation | 5,451 100 | 1.1 | 2 |
| Waste heat recovery No. 3 | 8,428,100 | 1.7 | 1-1/2 |
| Oxygen plant | 29,184,000 | 5.8 | 2 |
| Sulfur recovery | 6,887.000 | 1 4 | 3 |
| Waste water treatment | 14,013.500 | 2.8 | 2 |
| Flue gas processing | 20,141,200 | 4 0 | 2 |
| Drying | 910,600 | .2 | 1 |
| Steam and power plant | 41,093 000 | 8.2 | 4 |
| Plant facilities | 24.547.400 | 4 9 | 6 |
| Plant utilities | 35,184,700 | 7.1 | 4 |
| Total construction | 387,031,400 | 77.5 | 50 |
| Initial catalyst requirements | 7,455 100 | 1.5 | |
| Total plant cost (insurance and tax bases) | 394,486,500 | 79.0 | |
| Interest during construction | 59,173 000 | 11.9 | |
| Subtotal for depreciation | 453,659,500 | 90.9 | |
| Working capital | 45,366 000 | 9.1 | |
| Total investment | 499,025,500 | 100.0 | |

Source: (33,p.29)



TABLE 2.9

**Estimated Annual Operating Cost for HYGAS Plant**

| | | Unit Cost | Total | Percent |
|---|---|---|---|---|
| Direct cost: | | | | |
| Raw materials and utilities: | | | | |
| Coal | 665.7 tph x 7,920 hr/yr x $11/ton | $57,995,800 | | |
| Raw water | 1,200 M gph x 7,920 hr/yr x $0.15/M gal | 1,425,600 | | |
| Catalyst and chemicals | | 3.408,600 | | |
| Methane | 77.5 M scfh x 7,920 hr/yr x $0.75/M scf | 46.400 | | |
| | | | $62,290,400 | 51.3 |
| Direct labor: | | | | |
| 1,200 man-hr/day | $6/man-hr x 365 day/yr | 2,628.000 | | |
| Supervision | 15% of labor | 394,2000 | | |
| | | | 3,022,200 | 2 5 |
| Plant maintenance: | | | | |
| 467 men | $15.000/yr | 7,305 00 | | |
| Supervision | 20% of maintenance labor | 1,461 000 | | |
| Material and contract | 150% of maintenance labor | 10,957,500 | | |
| | | | 19,723,500 | 16 0 |
| Payroll overhead | 30% of payroll | | 3,530,500 | 2.9 |
| Operating supplies | 20% of plant maintenance | | 3,944,700 | 3.2 |
| Total direct cost | | | 93,517,300 | 75.9 |
| Indirect cost (administration and general overhead) | | | | |
| | 40% labor, maintenance, and supplies | | 10,676,200 | 8 7 |
| Fixed cost: | | | | |
| Taxes and insurance | 2% of plant cost | | 7.889,700 | 6.4 |
| Depreciation | 5% of subtotal for depreciation | | 22,683 000 | 17 9 |
| Total operating cost, before credits | | | 134,766,200 | 108.9 |
| Credits: | | | | |
| Sulfur | 185.04 tpd x 330 day/yr x $25/ton | | 1,526,600 | 1.2 |
| Ammonia | 133.42 tpd x 330 day/yr x $60/ton | | 2,641,700 | 2.1 |
| B-T-X | 83,271 gpd x 330 day/yr x $0.25/gal | | 6,869,900 | 5.6 |
| Operating cost, after credits | | | 123,728,000 | 100.0 |

Coal at $13 per ton, operating cost = $134,272,700.
Coal at $15 per ton, operating cost = $144 817,400
Source: (33,p.10)

TABLE 2.10

A Process Data Comparison of Normal and Slagging LURGI Gasifiers

| Item | Units | Fixed-bed Processes Process 11 LURGI Dry-Ash | Processes 18,19,20 LURGI Slagging |
|---|---|---|---|
| Coal (4.7% Moisture) | ton/day | 12,860 | 13,410 |
| Oxygen | ton/day | 5,700 | 8,050 |
| Cooling Water Circulated | gpm | 263,000 | 356,000 |
| Boiler Feed Water Makeup | gpm | 5,300 | 3,900 |
| Fired Boiler Steam Production | M lb/hr | 1,470 | 693 |
| Total Steam Production | M lb/hr | 4,765 | 3,103 |
| Dry Gas to CO Shift | MM scfd | 310 | 785 |
| Dry Raw Synthesis Gas | MM scfd | 883 | 802 |
| Methane by Synthesis | MM scfd | 133 | 166 |
| $CO_2$ Removal | ton/day | 16,050 | 18,500 |
| Sulfur | ton/yr | 49,000 | 59,000 |
| Ammonium Sulfate | ton/yr | 187,500 | 187 500 |
| Raw Phenols | ton/yr | 20,500 | 15,200 |
| Char | ton/day | --- | --- |
| Overall Efficiency BTU in Gas/BTU in coal | Percent | 66.9 | 64.1 |

Source: (10,p.74)



TABLE 2.11

Raw Gas Composition for Synthane Process

| Components | Mol % High-BTU | Low-BTU |
|---|---|---|
| CO | 10.5 | 5.4 |
| $CO_2$ | 18.2 | 14.9 |
| $H_2$ | 17.5 | 14.3 |
| $H_2O$ | 37.1 | 23.8 |
| $CH_4$ | 15.4 | 7.0 |
| $C_2H_6$ | 0.5 | 0.3 |
| $C_3H_6$ | | 0.1 |
| $H_2S$ | 0.3 | 0.3 |
| $N_2$ | 0.5 | 33.9 |
| Total | 100.0 | 100.0 |
| Higher heating value (dry basis), BTU/scf | 405 | 186 |

Source: (11,p.84)

TABLE 2.12

Total Estimated Capital Requirements and Operating Personnel for SYNTHANE Plant

| Unit | Cost | Percent | Estimated number of operators per shift |
|---|---|---|---|
| Coal preparation | $15,610,300 | 3.00 | 3 |
| Gasification | 83,062,300 | 15.98 | 7 |
| Dust removal | 17,282,200 | 3.32 | 2 |
| Shift conversion | 7,674,300 | 1.48 | 2 |
| Waste heat recovery | 13,263,200 | 2.55 | 1 |
| Purification | 53,950,800 | 10.37 | 4 |
| Methanation | 45,167,000 | 8.68 | 5 |
| Pipeline compression | 1,416,000 | .27 | 1 |
| Final drying | 1,169,500 | .22 | 1 |
| Oxygen plant | 37,050,000 | 7.12 | 2 |
| Sulfur recovery plant | 1,875,000 | .36 | 2 |
| Waste water treatment | 16,356,000 | 3.14 | 2 |
| Steam and powerplant | 52,667,100 | 10.12 | 6 |
| Plant facilities | 25,990,600 | 5.00 | 4 |
| Plant utilities | 37,253,400 | 7.16 | 4 |
| Toal construction | 409,787,900 | 78.77 | 46 |
| Initial catalyst requirements | 1,465,500 | .28 | |
| Total plant cost (insurance and tax bases) | 411,253,400 | 79.05 | |
| Interest during construction | 61,688,000 | 11.86 | |
| Subtotal for depreciation | 472,941,400 | 90.91 | |
| Working capital | 47,294,100 | 9.09 | |
| Total investment | 520,235,500 | 100.00 | |

Source: (32,p.11)

TABLE 2.13

Estimated Annual Operating Cost for SYNTHANE Plant

| | | Unit Cost | Total | Percent |
|---|---|---|---|---|
| Direct cost: | | | | |
| Raw materials and utilities: | | | | |
| Process coal | 593.7 tph x 7,920 hr/yr x $ 11/ton | $51,723,100 | | |
| Utility coal | 15.3 tph x 7,920 hr/yr x $11/ton | 1,332,900 | | |
| Raw water | 1,020 M gal/hr x 7,920 hr/yr x $0.15/M gal | 1,211,800 | | |
| Catalyst and chemicals | | 1,707,200 | $55,975,000 | 46.2 |
| Direct labor: | | | | |
| 1,104 man-hr/day | $6/man-hr x 365 day/yr | 2,417,800 | | |
| Supervision | 15% of labor | 362,700 | | |
| | | | 2,780,500 | 2.3 |
| Plant maintenance: | | | | |
| 508 men | $15,000/yr | 7,620,000 | | |
| Supervision | 20% of maintenance labor | 1,524,000 | | |
| Material and contracts | | 11,430,000 | | |
| | | | 20,574,000 | 17.0 |
| Payroll overhead | 30% of payroll | | 3,577,400 | 2.9 |
| Operating supplies | 20% of plant maintenance | | 4,114,800 | 3.4 |
| Total direct cost | | | 87,021,700 | 71.8 |
| Indirect cost (administration and general overhead) | 40% of labor, maintenance, and supplies | | 10,987,700 | 9.1 |
| Fixed cost: | | | | |
| Taxes and insurance | 2% of plant cost | | 8,225,100 | 6.8 |
| Depreciation | 5% of subtotal for depreciation | | 23,647,100 | 19.5 |
| Total operating cost, before credits | | | 129,881,600 | 107.2 |
| Credits: | | | | |
| Sulfur | 210.6 tpd x 330 day/yr x $25/ton | | 1,737,400 | 1.4 |
| Tar | 4,849 gph x 7,920 hr/yr x $0.10/gal | | 3,840,400 | 3.2 |
| Ammonia | 159.84 tpd x 330 day/yr x $60/ton | | 3,164,600 | 2.6 |
| Operating cost, after credit | | | 121,139,000 | 100.0 |

Operating cost, coal at $13 per ton - $130,785,600.
Operating cost, coal at $15 per ton - $140,432,200.

Source: (32,p.24)

TABLE 2.14

SASOL Liquefaction Products

| | | |
|---|---|---|
| Gasoline | Styrene | Ammonia |
| Fuel Oil | Higher Alcohols | Ammonia nitrate |
| Diesel Oil | Acetone | Ammonia Sulfate |
| Town Gas | Ketones | Crude Phenols |
| LPG | Xylol | Motor Benzole |
| Methanol | Naphtha | Benzene |
| Motor Alcohol | Waxes | Toluol |
| Ethanol | Paraffin | Road Tar Prime |
| Ethylene | Creosote | Electricity |
| Butadiene | Pitch | |

Source: (3,p.16)



TABLE 2.15

Product Composition of FISHER-TROPSCH

| | Composition (vol%) Fixed-bed Process | Fluid-bed Process |
|---|---|---|
| Liquified petroleum gas ($C_3$--$C_4$) | 5.6 | 7.7 |
| Petrol ($C_5$--$C_{11}$) | 33.4 | 72.3 |
| Middle oils (diesel, furnace, etc.) | 16.6 | 3.4 |
| Waxy oil or gatsch | 10.3 | 3.0 |
| Medium wax, mp 203°-206F. | 11.8 | |
| Hard wax, mp 203°-206°F.) | 11.8 | |
| Alcohols and ketones | 4.3 | 12.6 |
| Organic acids | traces | 1.0 |

| | Liquid-product Composition (vol%) Fixed-bed Process | | Fixed-bed Process | |
|---|---|---|---|---|
| Paraffins | 45 | 55 | 13 | 15 |
| Olefins | 50 | 40 | 70 | 60 |
| Aromatics | 0 | 0 | 5 | 15 |
| Alcohols | 5 | 5 | 6 | 5 |
| Carbonyls | traces | traces | 6 | 5 |

Source: (10,p.78)

TABLE 2.16

Yield Data for COED Pyrolysis of Illinois #6 Coal

| Net Process Yield | Wt.% of Dry Coal |
|---|---|
| 117 lb/ton | 59.1 |
| 1. 04 bbl/ton | 19.6 |
| 7. 1 gal/ton | 5.5 |
| 8.133 scf/ton | 15.8 |
| Total | 100.00 |

Source: (11,p.77)



TABLE 2.17

Typical SRC Yield

| | Wt % Raw Coal | Product Solvent Refined Coal |
|---|---|---|
| Carbon | 70.7 | 88.2 |
| Hydrogen | 4.7 | 5.2 |
| Nitrogen | 1.1 | 1.5 |
| Sulfur | 3.4 | 1.2 |
| Oxygen | 10.3 | 3.4 |
| Ash | 7.1 | 0.5 |
| Moisture | 2.7 | -- |
| | 100.0 | 100.0 |
| Volatile matter | 38.7 | 36.5 |
| Fixed carbon | 51.5 | 63.0 |
| Ash | 7.1 | 0.5 |
| Moisture | 2.7 | -- |
| | 100.0 | 100.0 |
| BTU/lb | 12,821 | 15.768 |

Source: (21,p.89)

TABLE 2.18

SRC Gas and Liquid Yields

| | |
|---|---|
| $C_1$-$C_4$ gas, scf† | 3,130 |
| $CH_4$, scf | 2,100 |
| $C_5$-350°F., gal<br>bbl | 32<br>0.762 |
| 350°-750°F. distillate, gal | 38 |
| bbl | 0.904 |
| Total liquid, gas<br>bbl | 70<br>1.666 |

*Per ton of solvent refined coal from hydroliquefaction reactor. †Approximate analysis of $C_1$-$C_4$ gas cut:

| | Vol % | BTU Value/cu ft |
|---|---|---|
| $CH_4$ | 67.0 | 680 |
| $C_2H_6$ | 19.3 | 340 |
| $C_3H_8$ | 10.0 | 260 |
| $C_4H_{10}$ | 3.7 | 120 |
| Total | 100.0 | 1,400 |

Source: (21,p.89)



TABLE 2.19

Total Estimated Capital Requirements and Operating Personnel for SRC Plant

| Unit | Cost | Percent | Estimated Number of Operators Per Shift |
|---|---|---|---|
| Coal preparation | $15,426,200 | 2.2 | 5 |
| Coal Slurrying and Pumping | 1,502,900 | .2 | 1/2 |
| Coal liquefaction and filtration | 151,361,000 | 21.6 | 5 |
| Dissolver acid gas removal | 55,007,000 | 7.5 | 1-1/2 |
| Coal liquefaction and product distillation | 8,096,600 | 1.1 | 1-1/2 |
| Fuel oil hydrogenation | 60,456,900 | 6.6 | 2-1/2 |
| Naphtha hydrogenation | 5,307,000 | .8 | 1/2 |
| Fuel gas sulfur removal | 4,423,900 | .6 | 1-1/2 |
| Gasification | 19,145,400 | 2.6 | 1-1/2 |
| Acid gas removal | 20,602,500 | 3.0 | 1-1/2 |
| Shift conversion | 16,496,000 | 2.4 | 1 |
| $CO_2$ removal | 9,751,000 | 1.4 | 1 |
| Methanation | 759,300 | .1 | 1 |
| Sulfur recovery | 4,200,000 | .6 | 1-1/2 |
| Oxygen plant | 25,000,000 | 3.6 | 1-1/2 |
| Product storage and slag removal | 15,994,700 | 2.3 | 1 |
| Steam and powerplant | 48,664,400 | 7.0 | 3 |
| Process waste water treatment | 3,513,300 | .5 | 1/2 |
| Plant facilities | 34,960,100 | 5.0 | 2 |
| Plant utilities | 50,109,400 | 7.1 | 2-1/2 |
| Total construction | 551,203,600 | 78.7 | 36 |
| Initial catalyst requirements | 2,669,300 | .4 | |
| Total plant cost (insurance and tax bases) | 553,872,900 | 79.1 | |
| Interest during construction | 83,080,900 | 11.8 | |
| Subtotal for depreciation | 636,953,800 | 90.9 | |
| Working capital | 63,695,400 | 9.1 | |
| Total investment | 700,649,200 | 100.0 | |

Source: (24,p.11)

TABLE 2.20

Estimated Annual Operating Cost for SRC Plant

| | | Unit cost | Total | Percent |
|---|---|---|---|---|
| Direct cost: | | | | |
| Raw materials and utilities: | | | | |
| Coal | $11/ton x 20,456 tpd @ 330 sd/yr | $74,255,200 | | |
| Raw water | 15 M gpm x $0.15/M gal x 1,440 min/day x 330 day/yr | 1,069,200 | | |
| Catalyst and chemicals | | 5,307,600 | | |
| | | | $80,632,000 | 47.4 |
| Direct Labor: | | | | |
| 864 man-hr/day | $6/man-hr x 365 day/yr | 1,892,200 | | |
| Supervision | 15% of labor | 283,800 | | |
| | | | 2,176,000 | 1.3 |
| Plant maintenance: | | | | |
| 684 men | $15,000/yr | 10,260,000 | | |
| Supervision | 20% of maintenance labor | 2,052,000 | | |
| Material | | 15,390,000 | | |
| | | | 27,702,000 | 16.2 |
| Payroll overhead | 30% of payroll | | 4,346,400 | 2.5 |
| Operating supplies | 20% of plant maintenance | | 5,540,400 | 3.2 |
| Total Direct Cost | | | 120,396,800 | 70.6 |
| Indirect cost (administration and general overhead) | 40% labor, maintenance, and supplies | | 14,167,400 | 8.3 |
| Fixed cost: | | | | |
| Taxes and insurance | 2% of plant cost | | 11,077,500 | 6.5 |
| Depreciation | 5% of subtotal for depreciation | | 31,847,700 | 18.7 |
| Total operating cost, before credits | | | 177,489,400 | 104.1 |
| Credits: | | | | |
| Sulfur | 634 tpd x $25/ton x 330 day/yr | | 5,230,500 | 3.1 |
| Power | 22,000 kW-hr/hr x $0.010/kW-hr x 330 day/yr x 24 hr day | | 1,742,400 | 1.0 |
| Operating cost, after credits | | | 170,516,500 | 100.0 |

Coal at $13 per ton, operating cost = $184,017,500.
Coal at $15 per ton, operating cost = 197,518,500.
Source: (24,p.34)



TABLE 2.21

H-COAL Products from Illinois No 6 Coal

| IBP cuts, °F. | Vol % | Gravity °API |
|---|---|---|
| IBP - 400 | 42.18 | 44.6 |
| 400-650 | 41.51 | 17.3 |
| 650-975 | 16.31 | 5.0 |
| Total | 100.00 | 25.2 |

Source: (11,p.79)

TABLE 2.22

Total Estimated Capital Requirements
and Operating Personnel for H-COAL Plant

| Unit | Cost | Percent | Estimated Number of Operators Per Shift |
|---|---|---|---|
| Coal preparation | $28,442,800 | 4.3 | 8 |
| Hydrogenation | 215,615,100 | 32.7 | 8 |
| Refinery gas cleanup | 15,941,200 | 2.4 | 1 |
| Oxygen plant | 31,000,000 | 4.7 | 2 |
| Hydrogen production | 61,289,000 | 9.3 | 6 |
| Hydrogen compression | 18,489,600 | 2.8 | 3 |
| Ammonia and hydrogen sulfide removal | 2,688,200 | 0.4 | 2 |
| Sulfur recovery | 4,500,000 | 0.7 | 1 |
| Tankage | 21,835,100 | 3.3 | 2 |
| Steam and powerplant | 38,332,100 | 5.8 | 3 |
| Plant facilities | 32,860,000 | 5.0 | 5 |
| Plant utilities | 47,099,300 | 7.2 | 3 |
| Total construction | 518,092,400 | 78.6 | 44 |
| Initial catalyst requirements | 3,086,900 | 0.5 | |
| Total plant cost (insurance and tax bases) | 521,179,300 | 79.1 | |
| Interest during construction | 78,176,900 | 11.8 | |
| Subtotal for depreciation | 599,356,200 | 90.9 | |
| Working capital | 59,935,600 | 9.1 | |
| Total investment | 659,291,800 | 100.0 | |

Source: (27,p.12)



TABLE 2.23

Estimated Annual Operating Cost
for H-COAL Plant

| | | Unit Cost | Total | Percent |
|---|---|---|---|---|
| Direct cost: | | | | |
| Raw materials and utilities: | | | | |
| Coal | 23,512 ton/day x 330 day/yr x $11/ton | $85,348,600 | | |
| Water | 660 Mgph x 24 hr/day x 330 day/yr x $0.15/M gal | 784,100 | | |
| Catalyst and chemicals | | 13,755,200 | | |
| | | | $99,887,900 | 54.4 |
| Direct labor: | | | | |
| 1,056 man-hr/day | $6/man-hr x 365 day/yr | 2,312,600 | | |
| Supervision | 15% of labor | 346,900 | | |
| | | | 2,659,500 | 1.4 |
| Plant maintenance: | | | | |
| 643 men | $15,000/yr | 9,645,000 | | |
| Supervision | 20% of maintenance labor | 1,929,000 | | |
| Material and contracts | | 14,467,500 | | |
| | | | 26,041,500 | 14.2 |
| Payroll overhead | 30% of payroll | | 4,270,000 | 2.3 |
| Operating supplies | 20% of plant maintenance | | 5,208,300 | 2.8 |
| Total direct cost | | | 133,417,900 | 75.1 |
| Indirect cost (administration and general overhead) | 40% labor, maintenance, and supplies | | 13,563,700 | 7.4 |
| Fixed cost: | | | | |
| Taxes and insurance | 2% of plant cost | | 10,423,600 | 5.7 |
| Depreciation | 5% of subtotal for depreciation | | 29,967,800 | 16.3 |
| Total operating cost, before credits | | | 192,022,300 | 104.5 |
| Credits: | | | | |
| Ammonia | 120.4 short ton/day x 330 day/yr x $60/ton | | 2,383,900 | 1.3 |
| Sulfur | 710.6 short ton/day x 330 day/yr x $25/ton | | 5,862,400 | 3.2 |
| Operating cost after credits | | | 183,776,000 | 100.0 |

Operating cost with coal at $13/ton = $199,293,900.
Operating cost with coal at $15/ton = $214,811,800.

Source: (27,p.23)

TABLE 2.24

**Plant Investment Costs for Second-Generation Low-Btu Gasificaion Schemes (Thousands of Dollars)**

| Plant Section | Case 1 | Case 2 | Case 3 |
|---|---|---|---|
| Coal Preparation | 2,850 | 2,228 | 2,228 |
| Oxygen Supplies | 7,064 | 5,992 | 1,774 |
| Gasification | 5,426 | 7,016 | 7,216 |
| Compression, Gas Expansion | 1,706 | 1,758 | 3,350 |
| Desulfurization and Dehydration | 889 | 904 | 1,622 |
| Sulfur Recovery | 1,332 | 2,701 | 1,780 |
| Interconnecting Piping | 1,825 | 1,353 | 1,492 |
| Utilities | 478 | 565 | 486 |
| Direct Field Cost | 21,994 | 21,941 | 20,372 |
| Distributable Field Cost | 2,195 | 1,352 | 1,628 |
| Total Field Cost | 24,189 | 23,293 | 21,000 |
| Engineering, Home Office, and Fee | 2,661 | 2,562 | 2,420 |
| Total Construction Costs | 26,850 | 25,855 | 24,420 |
| Startup Costs | 2,950 | 2,844 | 2,686 |
| Total Capital Costs | 29,804 | 28,699 | 27,106 |
| Basis | | | |
| Blast Mode | $O_2$/steam | $O_2$/steam | Air/steam |
| Gasifier Type | Entrained solids | Fluidized solids | Fluidized solids |
| Sulfur Content in Coal, % | 2 | 2 | 2 |
| Sulfur Emission in Gases, e ($SO_2$) lb $SO_2$ per $10^6$ Btu HHV of coal | 1.2 | 1.2 | 1.2 |
| Gas Treatment Pressure, optimum, psia | 150 | 150 | 150 |

Source: (47,p.5-3)



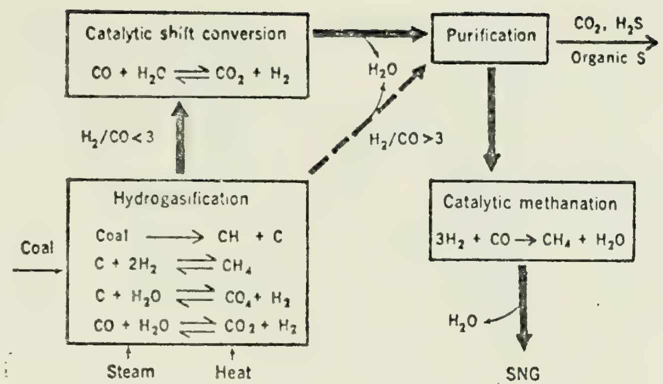
TABLE 2.25

**Annual Operating Costs for Second-Generation Low-Btu Gasification Schemes (Thousands of Dollars)**

| Cost Element | Case 1 | Case 2 | Case 3 |
|---|---|---|---|
| Coal @ $25/ton | 2,969 | 2,869 | 3,313 |
| Purchased electric Power @ $.030/kw-hr | 673 | 736 | 245 |
| Catalysts and Chemicals | 80 | 80 | 80 |
| Equipment, Supplies, Utilities | 80 | 80 | 80 |
| Operating Personnel | 528 | 528 | 528 |
| Maintenance Materials and Labor | 800 | 800 | 800 |
| Total Annual Operating Costs | 5,130 | 5,113 | 5,046 |
| Basis | | | |
| Blast Mode | $O_2$/steam | $O_2$ steam | Air/steam |
| Gasifier Type | Entrained solids | Fluidized solids | Fluidized solids |
| Sulfur content in coal, % | 2 | 2 | 2 |
| Sulfur Emission in Gases lb $SO_2$ per $10^6$ Btu HHV of Coal | 1.2 | 1.2 | 1.2 |
| Gas Treatment Pressure Optimum, psia | 150 | 150 | 150 |

Source: (47,p.5-2)

FIGURE 2.1

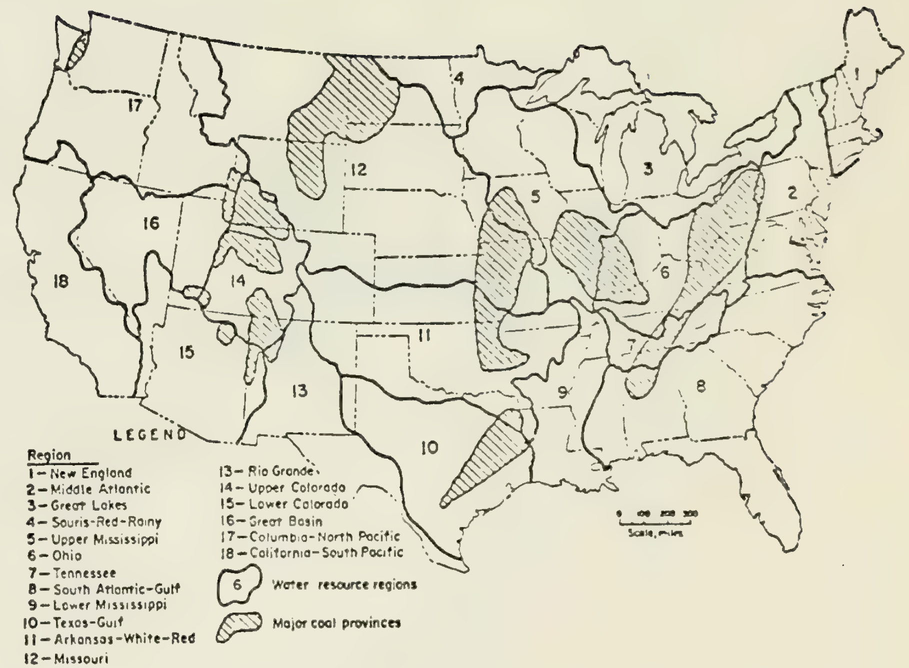



Generalized Gasification Flow Chart

Source: (5,p.45)



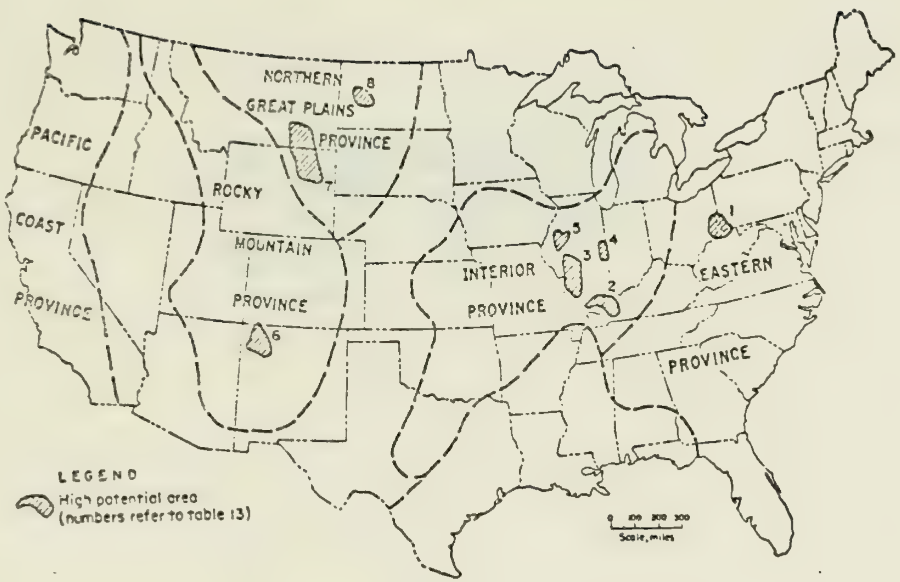
FIGURE 2.2




Major Water and Coal Resource Regions

Source: (6,p.27)

FIGURE 2.3

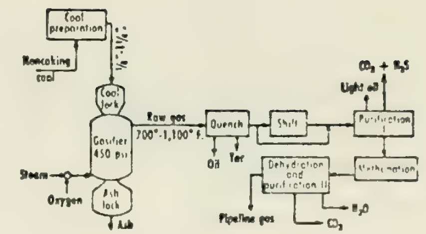



Areas of High Potential for Gasification Development

Source: (6,p.33)



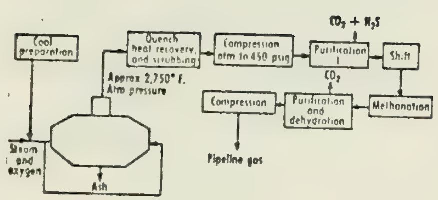
FIGURE 2.4




LURGI Process

Source: (11,p.80)

FIGURE 2.5

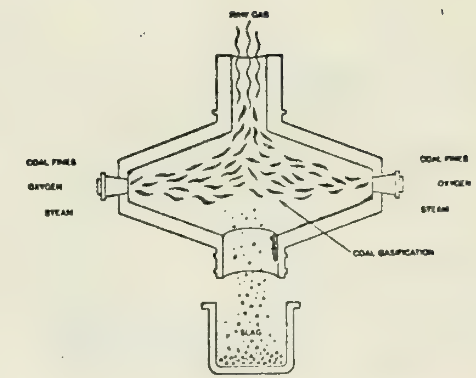



KOPPERS-TOTZEK Process

Source: (11,p.81)



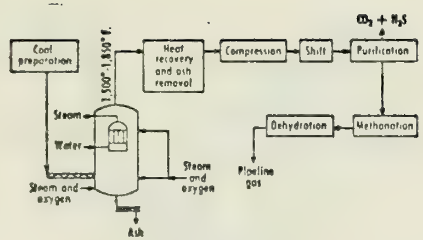
FIGURE 2.6




KOPPERS-TOTZEK Reactor

Source: (9,p.277)

FIGURE 2.7

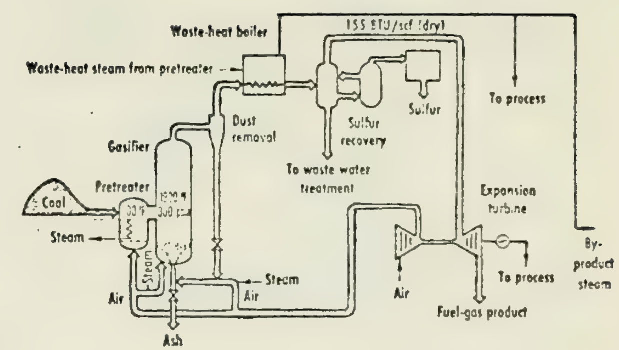



WINKLER Process

Source: (11,p.88)



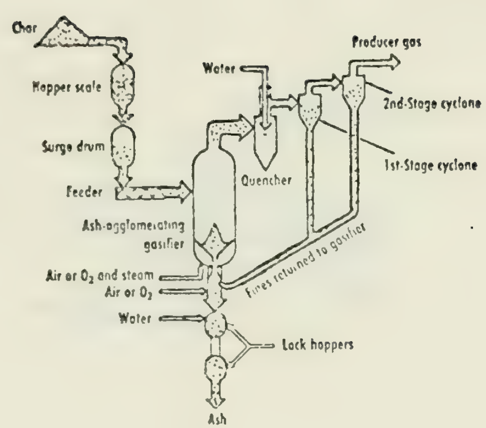
FIGURE 2.8




U-GAS Process

Source: (7,p.51)

FIGURE 2.9

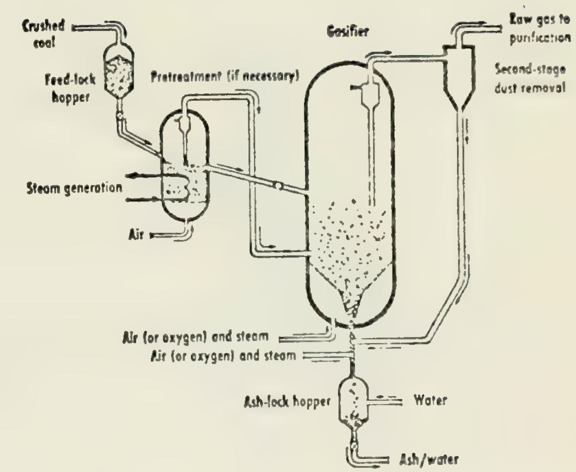



AAG Pilot Unit

Source: (7.p.53)



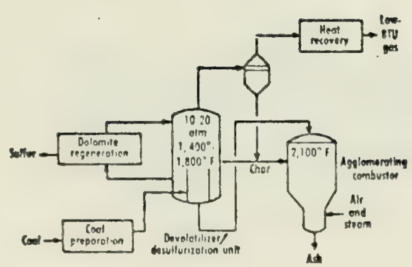
FIGURE 2.10




Modified U-Gas Process

Source (7.p.52)

FIGURE 2.11

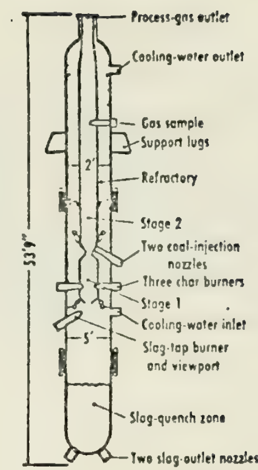



Westinghouse Coal Gasification Process

Source: (11,p.87)



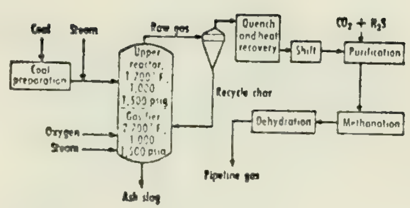
FIGURE 2.12




Pilot Plant BI-GAS Gasifier

Source: (13,p.56)

FIGURE 2.13

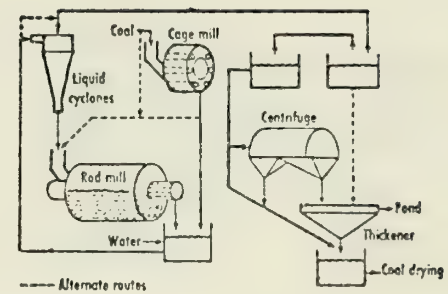



BI-GAS Process

Source: (11,p.75)



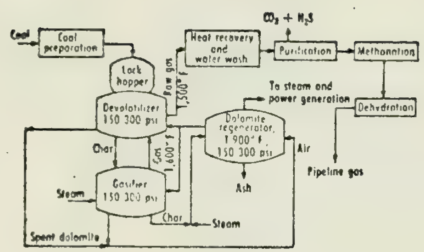
FIGURE 2.14




Coal Preparation Changes for BI-GAS Process

Source: (13,p.55)

FIGURE 2.15

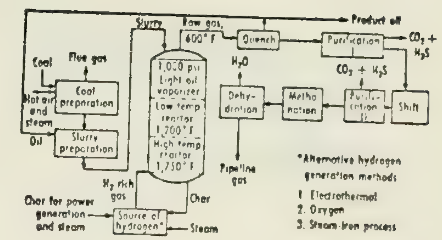



$CO_2$ Acceptor Process

Source: (11.p.76)



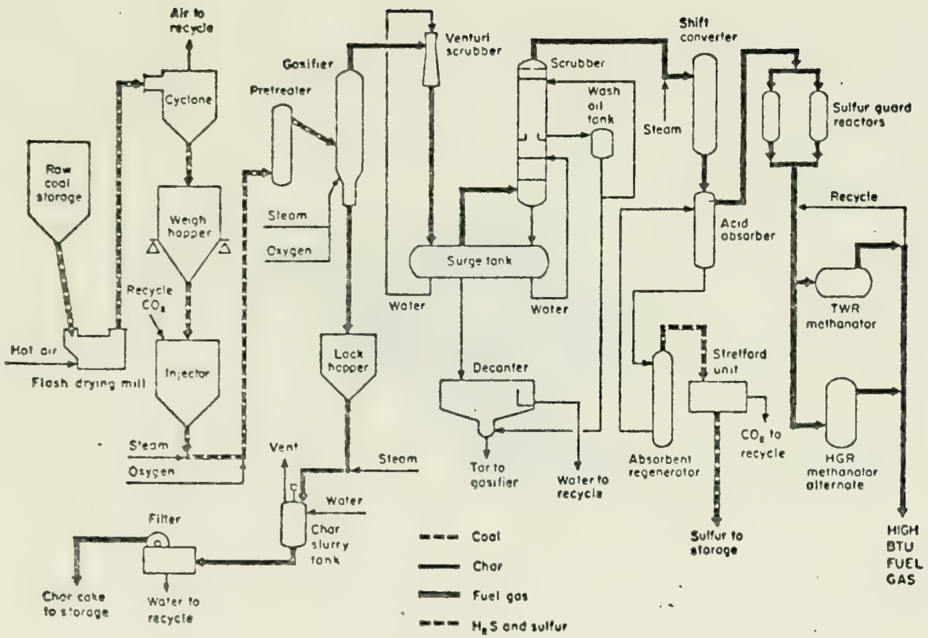
FIGURE 2.16




HYGAS Pilot Plant

Source: (15,p.75)

FIGURE 2.17

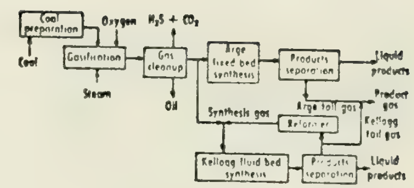



SYNTHANE Plant Flow

Source: (18,p.281)



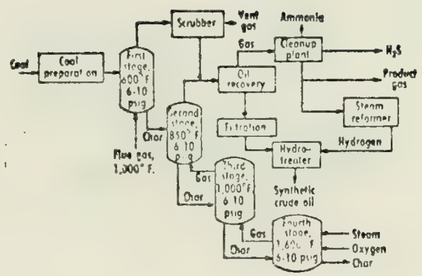
FIGURE 2.18




FISCHER-TROPSCH Process

Source: (11,p.77)

FIGURE 2.19

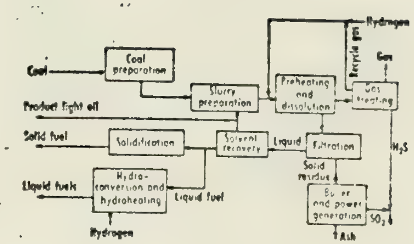



COED PROCESS

Source: (11,p.77)



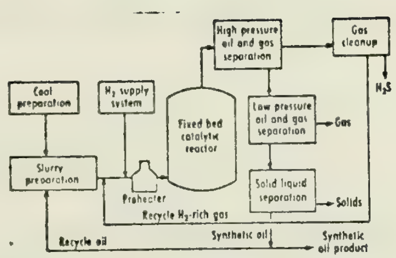
FIGURE 2.20




SRC PROCESS

Source: (11,p.83)

FIGURE 2.21

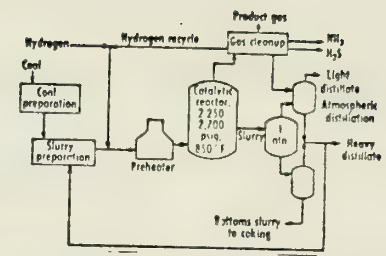



SYNTHOIL PROCESS

Source: (11,p.84)



FIGURE 2.22




H-COAL Process

Source: (11,p.79)

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# 3.0 COMBINED CYCLE

Among the routes considered for improved coal utilization are advanced power cycles. The primary advantage is increased fuel cycle efficiency, resulting in fewer pollutants for given levels of generated power [2]. The leading future alternatives are: 1) high temperature gas-steam turbine cycles. air-cooled turbine with hot-gas clean-up, 2) high-temperature gas-steam turbine cycle, ceramic turbine elements, utilizing pre-cleaned coal, and 3) advanced steam plants utilizing pressurized fluidized bed combustion (FBC) with in-bed sulfur removal (4). Major future improvements under this regime involve developing combined cycle configurations. Combined cycle processes are being studied by Westinghouse (7) and the Bureau of Mines (6). The Westinghouse system, in conjunction with its aforementioned Advanced Gasification System. appears to be the most advanced and economical process.

Figure 3.1 (1,p.2-8) illustrates a typical combined cycle process flow. A conventional combined cycle system involves a low-Btu gasification process followed by cleanup of the product gas before it is burned in a gas and steam turbine system (combined cycle). Note the waste heat recovery line from the gas turbine system to the steam boiler. This has been used effectively in many industrial applications (8). This overall

concept appears to have the lowest reported capital cost (1,p.2-20).

The parameters which will determine the future of this process are coal gasification advancement, environmental restrictions, and gas turbine technology. Future advantages include better resource utilization, lower natural gas requirements, reduced emissions, and higher thermal efficiency. Table 3.1 (1,p.2-2) compares some of the characteristics of a combined cycle system utilizing the COGAS process with those of a pulverized coal boiler with a tail-gas scrubber. Note that these are, as yet, only potential advantages. Extensive study has been conducted (4) for this process using Westinghouse's Advanced Gasification System (see Section 1). It is, however, not currently operational.

The advantages cited are dependent upon improvements in coal gasification techniques. This implies advances in second generation gasification techniques so that ORBES region coals can be used; eliminating the cost of excess coal transportation. That is, the cost-competitiveness of the entire system must justify increases in thermal efficiency. Improved thermal efficiency goes beyond process design for a combined cycle system. Improvements in gasification techniques not only result in better utilization of the coal, but create the possibility of higher temperatures for use in the system. However, this



requires the development of a high temperature gas clean-up system (primarily for $H_2S$) in order to retain this thermal potential. Gas turbine blades that can withstand increased temperatures are also required. Table 3.2 (1,p.2-23) lists the effects on efficiency of these improvements. Note that here the gasifier is not assumed to affect efficiency, but it will affect cost appreciably. The base case represents expected future improvements.

The thermal improvement capability of a combined cycle system is assumed to be an increase in efficiency from 36 to 44 percent (1). However, this assumes solution of all technological (and relevant economic) questions. This is required in order to offset the high investment costs. The time horizon for combined cycles is well into the future. For example, the discussion here assumes improved coal gasification techniques and a high temperature process in which the thermal potential is retained by the future development of a hot-gas cleanup system. Hot-gas cleanup is currently possible only for streams with very few tars and oils. In general, a commercial hot-gas cleanup process will not be available until at least 1985 (1,p.2-18), as the one closest to development is the rather expensive iron oxide process. A dolomite-based hot cleanup process (9) also seems favorable to alternative wet gas cleanup techniques, but this has

been proven only on a bench scale.

High temperature handling is desired so that the turbine gas inlet temperature will be higher. Table 3.3 (1,p.2-22) illustrates some desirable operating specifications. However, note that high turbine gas inlet temperatures are not seen to be feasible unitl 1995. This is due to the need to improve the metallurgical characteristics of the gas turbine blades in order to cope with these higher temperatures. At present, 2000$^{o}$F is the maximum allowable temperature. Improved blade materials (such as ceramics) are also needed in order to withstand possible damage from fines entrained in the hot fuel gas stream. Cyclones appear to be the particular removal system least affected by high pressure and temperature, but this rather inexpensive system has a particulate size limitation of about 5 microns or larger (1,p.2-16).

Clearly, many of the component problems in a combined cycle system place the earliest available date of an effective commercial process well into the 1990's. Studies are still required concerning instrumentation, multiphase flow characteristics, combustion characteristics, and erosion control [5]. This does not include the integration of these components into a single operating unit. In addition, the high temperature conditions increase $NO_x$ emission from the gas turbine; an increasingly important environmental consideration.



Studies (3) would also seem to indicate that hot gas cleanup systems with second generation gasification schemes may not offer the large thermal and economic incentives initially anticipated. Tables 3.4 (3,p.5-8) and 3.5 (3,p.5-9) list comparative capital costs for combined cycle systems at low and high termperatures, respectively. This analysis was done for certain reactor characteristics and gas cleanup systems, the characteristics of which are defined in Table 3.6 (3.p.4-7). A similar comparison of operating costs is given in Tables 3.7 (3,p.5-10) and 3.8 (3,p.5-11).

Because there are characteristic start-up and turn-down problems it appears that this system would have to be used as a base-load operation. It does not now appear that the tradeoff between potential operating advantages and subsequent required improved technology of the high-temperature combined cycle system will allow this process to be commercially effective within the desired ORBES time horizon. This goes beyond the dependence on the development of a second generation gasification technique and subsequent environmental control process. Table 3.9 (10,p.35) compares the status of various advanced power systems.

TABLE 3.1

**Potential Environmental Impact of 1000 MW Power Plant**

| | Pulverized Coal Boiler With Stack Gas Scrubber | Gasification-Combined Cycle Power Plant |
|---|---|---|
| Coal Consumption: LB/KWH | 0.82 | 0.70 |
| Limestone Required: LB/KWH | 0.15-0.20 | 0 |
| $NO_x$ Emissions: PPM | 500 | 400 |
| $SO_2$ Emission: PPM | 200 | 10-200 |
| Particulate Emissions: GR/SCF | 0.01 | Comparable |
| Cooling Tower Heat Rejection: BTU/KWH | 4,200 | 2,500 |
| Makeup Water: GAL/KWH | 0.6-0.65 | 0.4-0.45 |
| Disposal Land Required: Acres/1,000 MW | 1,200-2,400 | 200 |
| Projected Maximum Efficiency (1990) | 36% | 44% |

Source: (1,p.2-2)



TABLE 3.2

**Typical Efficiency Considerations and Effects**

| | Base Case | Revised Design Condition | Change In Efficiency % | |
|---|---|---|---|---|
| 1. | Turbine Inlet Temperature | 2800°F | 2000°F | -8.0% |
| 2. | Gas Cleanup Temperature | 100°F | 800°F | +1.0% |
| 3. | Gasifier Pressure | 300 psi | atmospheric | -1.0% |
| 4. | Gasifier Type | Second Generation | First Generation | 0% |
| 5. | Air vs. Oxygen Blown | Air | Oxygen | -1.0% |

Source: (1,p.2-23)

TABLE 3.3

Combined Cycle Specifications

| | |
|---|---|
| Steam Plant: | 2400 psi, 1000°F, 1000°F (reheat), heat rate = 9000 Btu/kwhr, unfired waste heat recovery system |
| Gas Turbine: | Inlet temperature 2000°F to 2800°F (1975) (1995) |
| Combined Cycle: | Capacity Factor: base load = 0.70<br>Intermediate Load = 0.45<br>Power Split: 58% gas turbine, 42% steam turbine |

Source: (1,p.2-22)

TABLE 3.4

Summary of Estimated Component
Capital Costs 1,950°F Gas Turbine Inlet Temperature

| Case | | MXL | MXH | MAL | MAH | MSL | MSH | EXL | EXH | EAL | EAH |
|---|---|---|---|---|---|---|---|---|---|---|---|
| Output Mw (100%) | | 1,000 | 1,000 | 1,000 | 1,000 | 1,000 | 1,000 | 1,000 | 1,000 | 1,000 | 1,000 |
| Coal M Lb/Hr | | 950.4 | 787.1 | 898.6 | 753.8 | 764.4 | 743.4 | 768.0 | 762.9 | 734.1 | 725.7 |
| Coal Preparation | $MM | 14.169 | 11.751 | 13.415 | 11.254 | 11.412 | 11.038 | 11.764 | 11.369 | 10.960 | 10.834 |
| Gasification | $MM | 69.932 | 57.928 | 66.121 | 55.471 | 25.904 | 25.519 | 23.358 | 25.000 | 28.555 | 32.254 |
| Gas Cooling/Heating | $MM | 53.793 | 1.543 | 34.275 | 5.166 | 7.289 | 4.194 | 24.436 | 6.065 | 28.387 | 7.400 |
| Oxygen Unit | $MM | 69.644 | 63.934 | --- | --- | 56.322 | 53.719 | 70.958 | 74.704 | --- | --- |
| Air Compression | $MM | --- | --- | 28.343 | 23.776 | --- | --- | --- | --- | 28.772 | 28.391 |
| Water Treatment | $MM | 79.836 | --- | 75.482 | --- | 26.602 | --- | 7.943 | --- | 7.400 | --- |
| Cold Purification (1) | $MM | 37.840 | --- | 47.338 | --- | 26.753 | --- | 31.373 | --- | 38.003 | --- |
| Hot Purification | $MM | --- | 73.021 | --- | 72.968 | --- | 52.349 | --- | 56.146 | --- | 61.366 |
| Combined Cycle | $MM | 176.900 | 200.000 | 178.146 | 200.000 | 200.000 | 200.000 | 200.000 | 200.000 | 200.000 | 200.000 |
| Offsites | $MM | 109.204 | 70.480 | 97.389 | 63.182 | 61.383 | 60.095 | 63.225 | 64.059 | 56.487 | 58.173 |
| Tar Fired Power Unit | $MM | 115.656 | --- | 109.369 | --- | --- | --- | --- | --- | --- | --- |
| Subtotal | $MM | 726.996 | 478.656 | 649.878 | 431.817 | 415.665 | 406.974 | 433.057 | 439.365 | 400.514 | 398.438 |
| Contingency | $MM | 109.050 | 71.798 | 97.480 | 64.772 | 62.350 | 61.046 | 64.959 | 65.905 | 60.077 | 59.766 |
| Plant Investment | $MM | 836.050 | 550.454 | 747.358 | 496.589 | 478.015 | 468.020 | 498.016 | 505.270 | 460.591 | 458.204 |
| Illinois Sales Tax | $MM | 20.901 | 13.761 | 18.684 | 12.415 | 11.950 | 11.701 | 12.450 | 12.632 | 11.515 | 11.455 |
| Preproduction Costs | $MM | 46.421 | 32.022 | 43.397 | 28.963 | 27.977 | 27.360 | 29.129 | 29.472 | 26.951 | 26.601 |
| Royalties | $MM | 20.901 | 13.761 | 18.684 | 12.415 | 11.950 | 11.701 | 12.450 | 12.632 | 11.515 | 11.455 |
| Catalyst & Chemicals | $MM | 4.160 | 2.752 | 3.737 | 2.483 | 2.390 | 2.340 | 2.490 | 2.526 | 2.303 | 2.291 |
| Construction Interest | $MM | 146.188 | 94.920 | 128.874 | 85.632 | 82.479 | 80.705 | 85.878 | 87.129 | 79.424 | 79.011 |
| Total Depreciable Capital | $MM | 1,074.621 | 707.672 | 960.734 | 638.516 | 614.712 | 601.847 | 640.414 | 649.660 | 592.299 | 589.218 |
| Working Capital | $MM | 42.948 | 30.859 | 39.168 | 28.520 | 28.057 | 27.390 | 29.102 | 28.945 | 26.995 | 26.781 |
| Total Capital | $MM | 1,117.569 | 738.531 | 999.902 | 667.036 | 642.769 | 629.236 | 669.516 | 678.606 | 619.294 | 615.999 |
| $/kw Plant Investment | | 836 | 550 | 747 | 497 | 478 | 468 | 498 | 505 | 461 | 458 |
| $/kw Total Capital | | 1,117 | 739 | 1,000 | 667 | 643 | 629 | 670 | 679 | 619 | 616 |

Source: (3,p.5-8)

TABLE 3.5

Summary of Estimated Component
Capital Costs 2,400°F Gas Turbine Inlet Temperature

| Case | | MXL | MXH | MAL | MAH | MSL | MSH | EXL | EXH | EAL | EAH |
|---|---|---|---|---|---|---|---|---|---|---|---|
| Output, Mw (100%) | | 1,000 | 1,000 | 1,000 | 1,000 | 1,000 | 1,000 | 1,000 | 1,000 | 1,000 | 1,000 |
| Coal, M Lb/Hr | | 661.8 | 699.5 | 809.9 | 677.2 | 704.9 | 687.4 | 737.9 | 710.1 | 683.2 | 671.5 |
| Coal Preparation | $MM | 12.665 | 10.443 | 12.008 | 10.110 | 10.523 | 10.262 | 11.017 | 10.601 | 10.199 | 10.024 |
| Gasification | $MM | 63.421 | 54.902 | 59.556 | 53.118 | 28.037 | 21.681 | 26.653 | 23.560 | 28.347 | 29.870 |
| Gas Cooling/Heating | $MM | 48.764 | 1.370 | 30.872 | 4.637 | 6.720 | 3.712 | 22.886 | 7.507 | 26.419 | 6.853 |
| Oxygen Unit | $MM | 63.146 | 56.814 | --- | --- | 51.935 | 49.674 | 66.449 | 69.537 | --- | --- |
| Air Compression | $MM | --- | --- | 25.540 | 21.359 | --- | --- | --- | --- | --- | --- |
| Water Treatment | $MM | 72.367 | --- | 68.017 | --- | 24.539 | --- | 7.438 | --- | 6.886 | --- |
| Cold Purification | $MM | 34.309 | --- | 42.656 | --- | 24.669 | --- | 29.379 | --- | 35.365 | --- |
| Hot Purification | $MM | --- | 65.445 | --- | 66.061 | --- | 48.783 | --- | 52.609 | --- | 57.172 |
| Combined Cycle | $MM | 179.000 | 200.000 | 180.300 | 200.000 | 200.000 | 200.000 | 200.000 | 200.000 | 200.000 | 200.000 |
| Offsites | $MM | 102.241 | 67.035 | 91.119 | 60.794 | 60.092 | 57.849 | 62.099 | 62.018 | 57.011 | 56.359 |
| Tar Fired Power Unit | $MM | 104.918 | --- | 98.539 | --- | --- | --- | --- | --- | --- | --- |
| Subtotal | $MM | 681.071 | 456.009 | 608.687 | 416.078 | 407.216 | 391.961 | 425.951 | 425.852 | 390.954 | 386.547 |
| Contingency | $MM | 102.160 | 68.401 | 91.300 | 62.412 | 61.082 | 58.794 | 63.893 | 63.878 | 58.643 | 57.982 |
| Plant Investment | $MM | 783.231 | 524.410 | 699.987 | 478.490 | 468.299 | 450.755 | 489.844 | 489.730 | 449.597 | 444.529 |
| Illinois Sales Tax | $MM | 19.581 | 13.110 | 17.497 | 11.962 | 11.707 | 11.269 | 12.246 | 12.243 | 11.240 | 11.113 |
| Preproduction Costs | $MM | 45.293 | 30.392 | 40.569 | 27.814 | 27.531 | 26.306 | 28.567 | 28.499 | 26.232 | 25.927 |
| Royalties | $MM | 19.581 | 13.110 | 17.497 | 11.962 | 11.707 | 11.269 | 12.246 | 12.243 | 11.240 | 11.113 |
| Catalyst & Chemicals | $MM | 3.916 | 2.622 | 3.500 | 2.392 | 2.341 | 2.254 | 2.449 | 2.449 | 2.248 | 2.223 |
| Construction Interest | $MM | 135.060 | 90.429 | 120.708 | 62.511 | 80.753 | 77.728 | 84.469 | 84.449 | 77.528 | 76.655 |
| Total Depreciable Capital | $MM | 1,006.662 | 674.074 | 899.756 | 615.132 | 602.338 | 579.580 | 629.821 | 629.613 | 578.086 | 571.560 |
| Working Capital | $MM | 39.781 | 28.594 | 36.175 | 26.700 | 26.787 | 25.920 | 28.030 | 27.588 | 25.818 | 25.461 |
| Total Capital Requirement | $MM | 1,046.443 | 702.668 | 935.933 | 641.832 | 629.125 | 605.500 | 657.851 | 657.201 | 603.904 | 597.020 |
| $/KW Plant Investment | | 783 | 524 | 700 | 478 | 468 | 451 | 490 | 490 | 450 | 445 |
| $/KW Total Capital | | 1,046 | 703 | 936 | 642 | 629 | 606 | 658 | 657 | 604 | 597 |

Source: (3,p.5-9)



TABLE 3.6

Nomenclature for Identification of Cases

| Gasifier | Type | Purification | Gas Turbine Inlet Temperature |
|---|---|---|---|
| M=Moving Bed (LURGI) | X=Oxygen Blown | L=Low Temp (e.g. Benfield) | 1,950 °F |
| E=Entrained Bed | A=Air Blown | H=High Temp (Morgantown Process) | 2,400 °F |
| | S=Slagging/ Oxygen Blown | | |

Example: Case MXL 1950 refers to a Lurgi movng bed gasifier, oxygen blown with low temperature fuel gas desulfurization supplying fuel to a gas turbine with an inlet temperature of 1,950°F.

Source: (3,p.4-7)

TABLE 3.7

Summary of Estimated Development Costs
and Time to Operation of Demonstration Plant

| System | Time to Operation of Demonstration Plant (years) | Millions of Mid-1975 Dollars: R&D Costs Through Pilot Plant Operation | Millions of Mid-1975 Dollars: Demonstration Plant Costs | Millions of Mid-1975 Dollars: Total | Millions of Escalated Dollars: R&D Costs Through Pilot Plant Operation | Millions of Escalated Dollars: Demonstration Plant Costs | Millions of Escalated Dollars: Total | Comments |
|---|---|---|---|---|---|---|---|---|
| Advanced Steam-AFB | 6 | --- | 129 | --- | --- | 183 | --- | Includes AFB and PFB for demonstration but not not development costs. |
| Advanced Steam-PFB | 12 | --- | 160 | --- | --- | 340 | --- | |
| Closed Cycle Gas Turbines-Helium Organic Bottoming | 15 | 200 | 647 | 847 | 288 | 1373 | 1661 | Includes AFB for pilot and demonstration plant but not development costs. |
| Potassium Topping Cycle | 16 | 308 | 286 | 594 | 572 | 701 | 1273 | Includes PFB for pilot and demonstration plant but not development costs. |
| Open Cycle Gas Turbine Combined-Water Cooled-Semiclean Liquid Fuel | 14 | 150 | 142 | 292 | 207 | 300 | 507 | Excludes fuel processing development costs. |
| Open Cycle Gas Turbine Lttu Gasifier | 11 | 46 | 137 | 183 | 55 | 233 | 288 | Includes gasifier for demonstration plant but not development costs.* |
| Open Cycle MHD | 19 | 561 | 893 | 1444 | 825 | 2249 | 3074 | |

Source: (10,p.35)

TABLE 3.8

Estimated Cost of Electrical Power for a 2,400°F
Gas Turbine Inlet Temperature and a 70 Percent
Load Factor

| Case | MXL | MXH | MAL | MAH | MSL | MSH | EXL | EXH | EAL | EAH |
|---|---|---|---|---|---|---|---|---|---|---|
| Output (70%), Mw | 700 | 700 | 700 | 700 | 700 | 700 | 700 | 700 | 700 | 700 |
| Heat Rate, Btu/kWh | 10,544 | 8,556 | 9,907 | 8,265 | 8,624 | 8,409 | 9,028 | 8,666 | 8,359 | 8,215 |
| Total Capital Requirement, $MM | 1,046.443 | 702.67 | 935.933 | 641.63 | 629.125 | 605.5 | 657.65 | 657.20 | 603.90 | 597.02 |
| Operating Charges, $MM/Yr | | | | | | | | | | |
| Coal at $1/MM Btu | 64.656 | 52.478 | 60.750 | 50.604 | 52.862 | 51.564 | 55.360 | 53.275 | 51.257 | 50.374 |
| Operating Labor | 6.616 | 5.768 | 7.700 | 5.263 | 5.151 | 4.956 | 5.366 | 5.367 | 4.946 | 4.890 |
| Catalyst and Chemicals | 0.546 | 0.367 | 0.500 | 0.335 | 0.326 | 0.316 | 0.343 | 0.343 | 0.315 | 0.311 |
| Maintenance Labor | 12.923 | 8.653 | 11.550 | 7.695 | 7.727 | 7.437 | 8.062 | 8.060 | 7.416 | 7.335 |
| Materials | 16.014 | 12.061 | 16.100 | 11.005 | 10.771 | 10.367 | 11.266 | 11.264 | 10.341 | 10.224 |
| Administrative and Support Labor | 6.462 | 4.326 | 5.775 | 3.946 | 3.863 | 3.719 | 4.041 | 4.040 | 3.709 | 3.667 |
| General and Admin. Expense | 15.665 | 10.466 | 14.000 | 9.570 | 9.366 | 9.015 | 9.797 | 9.795 | 8.992 | 8.890 |
| Ad Valorem Taxes and Insurance | 19.561 | 13.110 | 17.447 | 11.962 | 11.707 | 11.269 | 12.246 | 12.243 | 11.240 | 11.113 |
| Utilities | 0.159 | 0.225 | 0.062 | 0.140 | 0.035 | 0.028 | 0.043 | 0.062 | 0.023 | 0.016 |
| Ash Disposal | 0.254 | 0.206 | 0.236 | 0.199 | 0.208 | 0.202 | 0.217 | 0.209 | 0.201 | 0.198 |
| | | | | | | | | | | |
| Capital Charges, $MM/Yr | | | | | | | | | | |
| Depreciation | 40.266 | 26.963 | 35.990 | 24.605 | 24.094 | 23.183 | 25.193 | 25.165 | 23.123 | 22.862 |
| Avg. Bond Interest, 8% | 21.347 | 14.334 | 19.093 | 13.093 | 12.834 | 12.352 | 13.420 | 13.407 | 12.320 | 12.179 |
| Avg. Return on Equity, 12% | 32.021 | 21.502 | 28.640 | 19.640 | 19.251 | 18.528 | 20.130 | 20.110 | 18.479 | 18.269 |
| Avg. Income Tax, 52% | 34.679 | 23.286 | 31.017 | 21.270 | 20.849 | 20.066 | 21.801 | 21.760 | 20.013 | 19.785 |
| | | | | | | | | | | |
| Cost of Services | 275.191 | 193.769 | 248.932 | 179.730 | 174.066 | 173.005 | 167.328 | 165.160 | 172.376 | 170.115 |
| Cost of Power, Mills/kWh (Coal = $1/MM Btu) | 44.9 | 31.6 | 40.6 | 29.3 | 29.2 | 28.2 | 30.6 | 30.2 | 28.1 | 27.7 |
| Cost of Power, Mills/kWh (Coal = $2/MM Btu) | 55.7 | 40.4 | 50.7 | 37.7 | 36.0 | 36.8 | 39.8 | 39.1 | 36.7 | 36.1 |

Source: (3,p.5-11)

TABLE 3.9

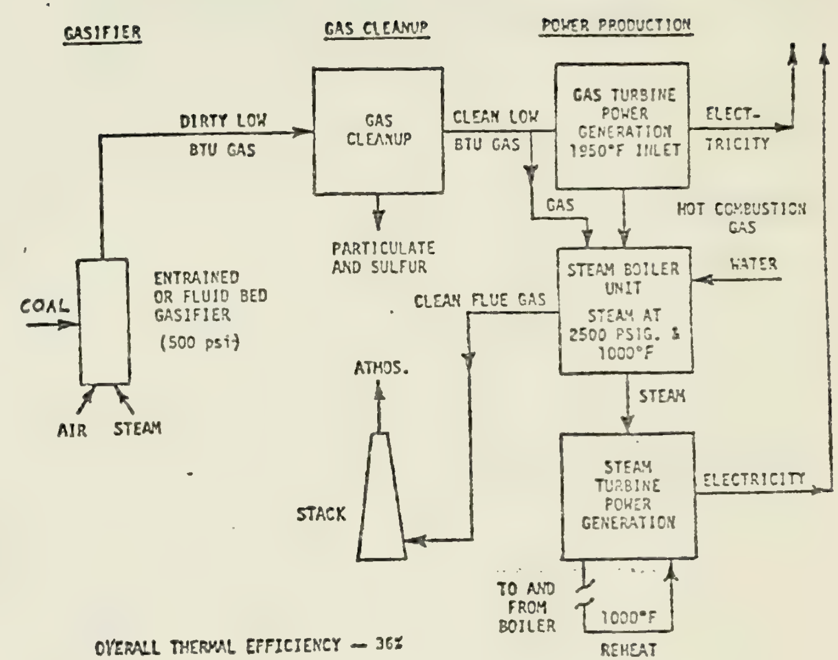
Summary of Estimated Development Costs
and Time to Operation of Demonstration Plant

| System | Time to Operation of Demonstration Plant (years) | Millions of Mid-1975 Dollars: R&D Costs Through Pilot Plant Operation | Millions of Mid-1975 Dollars: Demonstration Plant Costs | Millions of Mid-1975 Dollars: Total | Millions of Escalated Dollars: R&D Costs Through Pilot Plant Operation | Millions of Escalated Dollars: Demonstration Plant Costs | Millions of Escalated Dollars: Total | Comments |
|---|---|---|---|---|---|---|---|---|
| Advanced Steam-AFB | 6 | --- | 129 | --- | --- | 183 | --- | Includes AFB and PFB for demonstration but not development costs. |
| Advanced Steam-PFB | 12 | --- | 160 | --- | --- | 340 | --- | |
| Closed Cycle Gas Turbines-Helium Organic Bottoming | 15 | 200 | 647 | 847 | 288 | 1373 | 1661 | Includes AFB for pilot and demonstration plant but not development costs. |
| Potassium Topping Cycle | 18 | 308 | 286 | 594 | 572 | 701 | 1273 | Includes PFB for pilot and demonstration plant but not development costs. |
| Open Cycle Gas Turbine Combined-Water Cooled-Semiclean Liquid Fuel | 14 | 150 | 142 | 292 | 207 | 300 | 507 | Excludes fuel processing development costs. |
| Open Cycle Gas Turbine LBtu Gasifier | 11 | 46 | 137 | 103 | 55 | 233 | 288 | Includes gasifier for demonstration plant but not development costs. |
| Open Cycle MHD | 19 | 561 | 693 | 1444 | 825 | 2249 | 3074 | |

Source: (10,p.35)



FIGURE 3.1




Conventional Combined Cycle Process

Source: (1.p.2-8)

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# 4.0 DIRECT COAL COMBUSTION

## 4.1 Flue Gas Desulfurization

This section considers the state-of-the-art of flue gas desulfurization (FGD) techniques. FGD is used almost exclusively by the electric power industry. Alternatives to an FGD system include coal precleaning, SRC, FBC, or combined cycle coal gasification. While some FGD systems are time-proven, none are without undesirable features: 1) cumbersome and costly equipment for relatively small $SO_2$ gas volumes, 2) corrosion and erosion problems, and 3) difficulty with power plant integration due to their chemical nature. Some of the methods are expensive, still need improvement, or simply cannot meet the standards set by the 1977 Clean Air Act Amendments.

While FGD systems remain the most desirable cleaning process, as measured by utility acceptance, possible improvements provide a basis for the segregation of these techniques. First generation systems can be termed throwaway, while second generation techniques are regenerative. These terms refer to how the waste products are handled; the former refers to the disposal of the $SO_2$ and its absorbent, while the latter indicates recycling the absorbent and possible utilization of the wastes in



the form of some usable product. Table 4.1 lists the leading processes under their respective headings. There are obviously many more processes than those listed. However, those processes listed are the most advanced. Improvements in these will more than likely produce improvements in others. Table 4.2 categorizes the relevant information for each process. A chart formulated by Stukel (4,p.11), was used for much of the information. The listing of Status Ratings (SR) and Feasability Ratings (FR) are defined in the same way as those used for the coal gasification and liquefaction processes.

FGD is the primary emission control process used at this time. However, this discussion is not an attempt to compare throwaway and regenerative techniques. Regeneration does have potential advantages. It also has higher costs. Rather, a discussion of each process is developed with an examination of the main problems and costs of these processes. As lime/limestone is the most widely used currently available technique, the major portion of the discussion deals with this process. The comparable merits and disadvantages of the most feasible new techniques comprise the remainder of this section.

### 4.1.1 Lime-Limestone Wet Scrubbing (1,2.3.4)

This process is the most effective and, hence, most widely used to date. Of the 31 operating systems, 26 use this technique (2.p.ii). The ratio will probably increase. The technique involves using a slurry of calcium oxide or calcium carbonate to absorb $SO_2$ in a wet scrubber, with the resulting calcium sulfite and sulfate disposed of as waste solids. A flow diagram of the process is shown in Figure 4.1 (1,p.337). The overall absorption reactions are shown below:

<u>LIME</u> $SO_{2(g)} + CaO_{(s)} + 1/2\ H_2O \text{--->} CaSO_3 \cdot 1/2\ H_2O_{(s)}$

<u>LIMESTONE</u> $SO_{2(g)} + CaCO_{3(s)} + 1/2\ H_2O \text{--->} CaSO_3 \cdot 1/2\ H_2)_{(s)} + CO_{2(g)}$

The main problems associated with this process are (2.p.ii): $SO_2$ removal efficiency, wet-dry interface deposition, scaling, mist elimination, gas reheat, and corrosion-erosion. Future tests are required to determine the trade-offs in dealing with these problems. Even after an effective solution, the inherent problems of waste disposal and transportation of lime or



limestone remain.

Process efficiency depends on the amount and type of lime or limestone used for scrubbing. Figure 4.2 (2,p.36) shows a large available deposits in the ORBES region. The type of scrubber contacts is also very important for efficient absorption. Characteristics of the more popular types are shown in Table 4.3 (1,p.351).

The main problem with this technique is sludge disposal. This is substantial in both weight and volume, requiring a large waste pond or landfill. Table 4.4 (1.p.348) lists the waste product composition for this process. The environmental effects for disposal alternatives are listed in Table 4.5 (5,p.34), with some comparative cost ranges given in Table 4.6 (5,p.32).

A breakdown of the mass and energy requirements are given in Tables 4.7 (1,p.344) and 4.8 (1,p 345), for a typical limestone and lime scrubbing operation, respectively. Representative investment and operating costs are given in Tables 4.9 (2,p.56) and 4.10 (2,p.58) respectively. Tables 4.11 (3,p.55) and 4.12 (3,p.56) also compare some typical costs to those of other processes. Possible improvements (2,p.61) in the process may be made by the use of mass transfer additives, forced oxidation to improve sludge dewatering and scaling reduction, and sparged scrubbing.

### 4.1.2 Double Alkali (2,3,4)

This technique is actually a hybrid of the throwaway and regenerative processes. Flue gas $SO_2$ is removed by contact with a soluble alkali salt (sodium, potassium, or ammonia). This solution is then treated with another alkali element (calcuim) which regenerates the salt and removes a calcium-sulfur solid. The sludge produced is similar to that in the lime-limestone process. This technique has the advantages of scale-free scrubber operation and high $SO_2$ removal. If operated properly, it should have no corrosion problems. Stukel (4,p.22) illustrates the process and lists the relevant overall reactions.

The major drawbacks at this time are the possiblity of water pollution if not handled properly, the occurence of scaling if not operated properly, and that the system requires an external energy source, which reduces overall plant efficiency. Another problem is that the present process can use only lime which, because it requires a kiln operation, is more expensive and less energy efficient than limestone. The future of this process seems to depend on the operating characteristics of a large scale plant. The capital and operating costs in Table 4.9 and 4.10 should be noted.



### 4.1.3 Sodium Scrubbing (1,2,3,4)

This technique involves the absorption of $SO_2$ in a sodium sulfite/bisulfite solution. A simplified process diagram, with reactions, is shown by Stukel (4,p.31). A detailed diagram illustrating the equipment is shown in Figure 4.3 (1,p.247). The spent scrubbing solution is processed to regenerate the scrubbing solution and separate the absorbed $SO_2$. The $SO_2$ is then processed into a final product for marketing. The obvious advantages are a reusable scrubber and a marketable product to reduce costs. A key advantage is that the scrubbing process is proven and well-defined.

The Wellman-Lord process uses the thermal regeneration technique for recovering the scrubbing solution. This simply involves reversing the absorption reaction by the addition of heat. The product is a concentrated $SO_2$ stream that can then be converted into elemental sulfur or sulfuric acid. The primary disadvantage is that this process may be very expensive. It is energy intensive, and a significant amount of sodium sulfate is produced which must be removed. The major drawback may be that currently natural gas is used as a reducing agent (for $SO_2$ conversion). Typical mass and energy requirements are given in Table 4.13 (1,p.256). At this time it seems the future of the

process lies in small industrial applications.

A process variation, developed by Atomics International, may be more promising. This is another sodium scrubbing operation, but sodium carbonate is used as the absorbent. The major advantage is that coal (or coke) may be used as the reducing agent for the solids. This yields sodium sulfide which is converted to $H_2S$ and, ultimately, elemental sulfur. Tests are still being done on the overall process.

### 4.1.4 Magnesia Scrubbing (1,2,3,4)

This technique is probably the simplest and most promising of the advanced processes. Its main advantages are the ease of absorbent regeneration, the ability to separate the power and chemical units, and the minimal effect on the power plant due to the elimination of a solids disposal problem. A magnesia-based slurry is used to absorb $SO_2$ to form magnesium-sulfur compounds. The scrubbing agent is then regenerated while the recovered $SO_2$ is processed to form liquid $SO_2$, sulfuric acid, or elemental sulfur. The primary absorbent reaction is:

$$Mg(OH)_2 + SO_2 \text{---}> Mg\ SO_3 + H_2O$$

The magnesium oxide is regenerated by thermal decomposition:



$$Mg + SO_3 \text{---}> MgO + SO_2$$

Figure 4.4 (1,p.311) illustrates a typical magnesia slurry scrubbing process. A magnesia scrubbing process can use a magnesia base slurry, solution, or solid, but the slurry system is the most advanced.

Current estimated costs make this system competitive, and there do not seem to be any undesirable products. However, due to the relatively low $SO_2$ concentration in the kiln offgas, the primary conversion process being used yields sulfuric acid rather than elemental sulfur. This acid is not as good a product as elemental sulfur as it is impractical to store for extended periods of time. In addition, the market is relatively small. This process is also somewhat energy intensive, and there are questions concerning the regenerative capabilities of magnesia.

However, the Chemico-Basic magnesia slurry $SO_2$ recovery system for control of sulfur oxides has been demonstrated successfully on an oil-fired boiler at Boston Edison, and has been tested on a coal-fired boiler at Potomac Electric [6]. The system showed the ability to remove 90% of the $SO_2$ present. Particulate control capability of the system was also demonstrated. Magnesia was regenerated and recycled successfully, and commercial grade sulfuric acid was produced and marketed. Long duration runs are still required for the system

because major shutdowns occured due to corrosion problems.

### 4.1.5 Comments

Currently the lime/limestone slurrying technique is the proven process. It will remain so until newer processes are refined. The magnesia-based system seems to be the most desirable regenerative technique. Future pollution regulations could dictate an even larger use, and development, of FGD systems. A possible enhanced advantage for regenerative techniques is related to a potentially massive sludge disposable problem which could be offset by the much smaller volumes of marketable or disposal sulfur. Other techniques, such as catalytic oxidation, have also shown desirable features, but at this point it seems they are still at least a decade away. In any case, the requirements of raw materials, water, energy and manpower, will be constant determinants. This includes requirements for scrubber back-up systems and maintenance. In fact, as indicated by future construction plans, it appears that the first generation lime/limestone scrubber will be the primary process for quite some time.

However, it must be kept in mind that just as EPA standards encouraged the use of FGD systems, the standards could also change to influence the position of the throwaway systems. That



is, EPA regulations imposed on throwaway landfills or ponds could make the system costs prohibitive. The desirability of the regenerative systems would then be enhanced. Refinement of the regenerative processes would then be accelerated. The by-product sulfur market would then compete with Frasch process sulfur.

## 4.2 Fluidized Bed Combustion

Fluidized bed coal combustion (7,8.9) is of major technological interest due to the advent of strict environmental restrictions on combustion processes and the increasing costs of clean-burning oil and gas. There are three regimes of sulfur removal, a primary criterion, for coal combustion:

1 Removal of sulfur before combustion-coal precleaning

2. Removal of sulfur oxides after combustion-flue gas desulfurization (FGD).

3. Removal of sulfur compounds during combustion-fluidized bed combustion (FBC)

This third is of concern here. Because coal precleaning will not have a significant commercial impact even with improving technology before 1985 (7 p.5-1), comparisons made here will be

between FBC and FGD systems.

Fluidized bed combustion involves a mass of particles undergoing combustion while held in suspension by an upward flow of air. This suspension is in contrast to other modes of coal feed: entrained and moving beds. Carbon particles in the flue gas are collected and recycled for further burning. The ideal particle size for the system is 1/4 to 3/4 inch maximum (6,p 5) because this eliminates the need for pulverization equipment. This size seems to burn quickly and completely. Desulfurization is accomplished by the use of limestone which is fed directly into the bed. The primary reaction for sulfur removal is shown below (6,p.4):

$$CaCO_3+SO_2+1/2O_2 \dashrightarrow CaSO_4+CO_2$$

The main components of an FBC unit operation are: 1) reaction vessel, 2) solids feed or flow control, 3) solids discharge, 4) dust separator, 5) instrumentation, and 6) gas supply (12)

The potential advantages of an FBC system over conventional combustion with FGD are offered below (7.p.5-2).

1. Lower coal preparation cost

2. The ability to use coals with high ash content

3. Higher heat transfer - less heat transfer surface required



4. Higher heat generation - smaller boilers

5. Reduced corrosion due to lower temperature

6. Lower $SO_2$ emission

7. Lower $NO_x$ emission

8. The possibility of passing the stack gas through a gas turbine for increased efficiency

However, additional work is required to determine the optimum characteristics for various aspects of the process (9). Scale-up is required to determine the optimum bed design for the system. This includes the use of limestone for $SO_2$ control. (Like most FGD systems, limestone supply and throwaway are additional cost considerations.) More research is needed in order to understand the heat transfer characteristics and to improve the overall combustion efficiency. Further study is also needed to determine the erosion limits of the materials used. Although at this time $NO_x$ and $SO_2$ standards are met, a reduction in these standards could become a factor, limiting the use of FBC systems. $NO_x$ emission could also increase with increased coal combustion (due to reaction with fuel-bound nitrogen) or from changes in

operating temperature.

There are two types of FBC systems - atmospheric and pressurized. An atmospheric system could be used for a convensional utility boiler steam cycle. A typical system is shown in Figure 4.5 (7,p.5-7). Quenching caused by the boiler tubes becomes an additional design consideration. This boiler would replace the conventional pulverized or cyclone fired boiler. Fly ash is removed by electrostatic precipitation, and air heaters for recovery are included, as well as most of the conventinal auxiliaries. Monongahela Power Company has installed and is testing (11,p.2) a 30-Mw atmospheric demonstration boiler at Rivesville, West Virginia (8,p.5). This is a Foster-Wheeler product.

Pressurized (approximately 10 atm.) FBC systems would be used primarily with a combined gas/steam turbine cycle. Here the combustor would be used as a supercharged boiler between the air compressor and the gas turbine. A typical system is shown in Figure 4.6 (7,p.5-8). Coal dust combustion is directly controlled by the amount of air supplied to the fuel. Some of the attractive features of a pressurized FBC system include (7,p.5-9):

1. Increased combustion rates due to lower temperature-less carbon carryover

2. Higher steam production with smaller bed size



increase

3. Lower $NO_x$ emissions

4. Coal to electric conversions as high as 45 percent-reduced capital investment over atmospheric FBC (possibly 20 percent)

One of the limitations of this system is the temperature. Above 1900°F there is a shift in the chemical equilibrium for sulfur removal and free $SO_2$ increases. In addition turbine blade materials cannot tolerate tempertures much greater than this. There are also design problems, because of excessive corrosion of boiler tubes due to particle motion and turbine blade damage due to particulates. These have been seen experimentally as damaging possibilities. A pilot plant using pressurized FBC is planned by the Curtiss - Wright Company at Woodridge, New Jersey (8,p.6). This 13-MW plant is scheduled for start-up in 1980.

In addition to the temperature limitations mentioned above, both types of FBC systems must maintain a temperature of at least 1400°F. Below this the combustion efficiency is too low, the $SO_2$ removal reaction is too slow, and the heat transfer advantages are lost. Thus the operating temperature range is quite restrictive. In addition turndown is, at most, 50 percent of full load (and usually 75 to 80 percent (7,p.5-5)), restricting its use for for variable load operation. The start-up of an FBC system is also quite complex, and will probably require some type

of external firing.

Despite the shortcomings mentioned above, a comparative cost analysis (7) was done for conventional combustion systems with FGD, and atmospheric and pressurized FBC systems. The cost estimates assume some resolution of operating problems, but at best are only of comparative interest. The results for capital and operating costs are shown in Table 4.14 (7,p.5-13) and Table 4.15 (7,p.5-14), respectively. However, 80% sulfur removal has been assumed for the FBC systems, which is now unacceptable. This would cause an increase in operating costs, if an increase in removal is possible at all.

If the design and operating problems mentioned above can be resolved, Tables 4.14 and 4.15 demonstrate a possible advantage for FBC systems. Howevever, at this time it appears that while atmospheric FBC systems might be applicable in specific situations within the near future, pressurized FBC systems will probably not be commercially available until the 1990's at best. A possible use for an atmospheric FBC system might be as a base load operation for a power plant, since startup and turndown seem to be major operational problems. Even here coal type (e.g. sulfur content) could be a limiting factor. Another possible utilization is industrial process heating, such as in a refinery or other cogeneration units:



> "But industry, where nearly one third of all U.S. energy is consumed, was not receiving attention. Primarily for this reason, the Energy Research and Development Administration (ERDA) issued contracts to develop FBC's for industrial uses.
>
> "An evalution of FBC's for refinery and chemical-plant fired process heaters is under way at Exxon Research and Engineering Co. the goal of Exxon's program is to make a complete technical and economic assessment of the FBC concept." (10,p.96)

Industrial utilization may be a more feasible long-term application for all FBC systems rather than competing with conventional utility techniques. Fluidized bed systems have already been proven for industrial processes involving chemical reactions, drying, heating, or cooling in various stages of product manufacture (12). Thus application of coal combustion in a fluidized state would seem to be a viable alternative for industries in or out of the ORBES region.

TABLE 4.1

| THROWAWAY | REGENERATIVE |
|---|---|
| Wet Lime/Limestone | Sodium Scrubbing - Thermal Regeneration |
| Double Akali | Sodium Scrubbing - Melt Reduction<br>Magnesia Scrubbing |



TABLE 4.2

Status of FGD Systems

| FR | SR | Process Name | Developed By | Process Remarks | Status Remarks | Fuel Type | Eff. |
|---|---|---|---|---|---|---|---|
| 1 | 1 | Limestone Scrubbing | --- | Throwaway Process; Sludge Disposal Problem | Commercially Available | Coal | 70-85 |
| 4 | 1 | Lime Scrubbing | --- | Throwaway Process; Sludge Disposal Problem | Commercially Available | Coal | 80-90 |
| 3 | 2 | Double Alkali Scrubbing | --- | Throwaway/Regenerative Process; Scaling and Corrosion Problems | Commercially Available (Construction) | Coal | 90-95 |
| 4 | 2 | Sodium Scrubbing Thermal Regeneration | Wellmen-Lord | Regenerative Process; Elemental Sulfur or Sulfuric Acid Product; Requires Natural Gas For Reduction | Commercially Available (Construction) | Coal | --- |
| 3 | 3 | Sodium Scrubbing Melt Reduction | Atomics International | Regenerative Process Uses Coal as Reducing Agent; Elemental Sulfur Product | DEMO. (TEST) | Coal | 80-90 |
| 2 | 3 | Magnesia Scrubbing | EPA, Philadelphia Electric | Regenerative Process; Sulfuric Acid Product, Simple Process | DEMO. (TEST) | Oil | 90 |

TABLE 4.3

COMPARISON OF SCRUBBER TYPES FOR A LIMESTONE WET SCRUBBING SYSTEM

| | Scrubber Type | | | | |
|---|---|---|---|---|---|
| Parameter | Marble Bed | TCA | Venturi | Grid Tower | Spray Tower |
| $SO_2$ Removal Efficiency | Good | Good | Fair | Good | Good |
| Particulate Removal Efficiency | Good | Good | Excellent | Good | Fair |
| Typical L/G (gal/1000 scf) for $SO_2$ Removal | 40-70 | 50-85 | 20-50 | 50-100 | 70-110 |
| Gas Side Pressure Drop (in $H_2O$) for L/G Above | 8-12 | 6-12 | 8-20 | 1-7 | 1-3 |
| Gas Velocity (ft/sec) | 3-8 | 6-11 | 125-300 | 6-11 | 5-25 |
| Dissolution of Solids | Good | Fair | Poor | Fair | Poor |
| Resistance to Solids Plugging | Fair | Good | Excellent | Fair | Excellent |

Source: (1.p.351)



TABLE 4.4

TYPICAL COMPOSITIONS AND QUANTITIES OF WASTE PRODUCT FROM LIME/LIMESTONE FGD SYSTEMS

| Lime Waste (dry) | Weight Percent | Production (tons/yr) | Assumed Packing Volume ($ft^3$/ton) of waste) | Approximate Volume Required For Storage in 30 Years (acre-feet) |
|---|---|---|---|---|
| $CaSO_3 \cdot 1/2H_2O$ | 60 | 85,741 | | |
| $CaSO_4 \cdot 2H_2O$ | 27 | 38,099 | | |
| $CaCO_3$ | 11 | 14,857 | | |
| Interts | 2 | 3.051 | | |
| TOTAL | 100 | 141,748 | 22 | 2.148 |
| Lime Waste Sludge (wet, 50% solids) | | 283,496 | 45 | 4,393 |
| Limestone Waste (dry) | | | | |
| $CaSO_3 \cdot 1/2H_2O$ | 53 | 85,741 | | |
| $CaSO_4 \cdot 2H_2O$ | 24 | 38,099 | | |
| $CaSO_3$ | 15 | 23.835 | | |
| Inerts | 8 | 12,493 | | |
| TOTAL | 100 | 160,168 | 22 | 2,427 |
| Limestone Waste (wet, 50% solid) | | 320.336 | 45 | 4.964 |

Assumptions:

Coal: 3.5% S
Plant: 500 Mw, 5,260 hr/yr, 375.000 lb/hr coal
Scrubber: 90% $SO_2$ removal
86% utilization of CaO
79% utilization of $CaCO_3$

Souuce: (1,p.348)

TABLE 4.5

ENVIRONMENTAL EFFECTS OF DISPOSAL ALTERNATIVES

| Type of Disposal | Condition of Waste | Primary Drainage | Environmental Effects: Seepage | Runoff | Land Reuse |
|---|---|---|---|---|---|
| Pond | Untreated[a] | Supernate | Yes | No | No |
| | or chemically fix[b] | Supernate | Yes | No | Yes |
| Basin | Untreated[a] or conditioned[c] | Underdrainage | Yes | No | No |
| Landfill | Conditioned[c] or chemically fixed[b] | Runoff | Yes | Yes | Yes |

[a]Untreated waste refers to FGD sludges as emitted from primary or secondary dewatering equipment.

[b]Chemically fixed sludges refer to the waste treated by one of several commercial processes that make these wastes suitable for landfill disposal.

[c]Conditioned waste refers to sludge treated by techniques other than chemical fixation and includes oxidation to gypsum and de-watering by mixing with dry fly ash or other agents that allow the material to be handled in a manner similar to that for soils.

Source: (5,p.34)



TABLE 4.6

SLUDGE DISPOSAL COST RANGES
(UNTREATED AND CHEMICALLY TREATED LANDFILLS,
1000-MW STATION.
50 PERCENT LOAD FACTOR.
30-YEAR AVERAGE.
JANUARY 1976 DOLLARS)

| Disposal Method | Base Material | S/Ton Sludge[a,b] (Dry) | \$/Ton Coal[b] | Mills/kWh[b,c,d] |
|---|---|---|---|---|
| Untreated Pond Pond | Natural Clay[e] Liner[f] | 3.50<br>5.70-7.80 | 1.00<br>1.60-2.20 | 0.4<br>0.7-1.0 |
| Chemically Treated[g] | Indigenous Soil | 7.30-11.40 | 2.10-3.20 | 0.9-1.4 |

[a]510,000 short tons/year average (dry basis) including fly ash.

[b]Coal burned at rate of 0.88 lb/kWh. 3% sulfur, 12% ash, 85% $SO_2$ removal, 1.2 $CaCO_3/SO_2$ mole ratio.

[c]Land costs at \$1000/acre are included (equivalent to \$0.25/ton sludge. dry).

[d]Disposal within 5 miles of power plant.

[e]Assumes coefficient of permeability of clay is $1 \times 10^{-6}$ cm/sec or better.

[f]Ponding costs cover range based on low-to-high material costs. i.e., PVC-20 (low) to Hypalon-30(high).

[g]Chemical fixation costs vary, depending on characteristics of the waste and the disposal process chosen.

Source: (5,p.32)

TABLE 4.7

Raw Material and Energy Consumption for Limestone Scrubbing

Basis: 500 Mw, 3.5 Percent Sulfur Coal, 5,260 hr/yr

| Process Area | Limestone tons/hr | Steam MM Btu/hr | Water M gal/hr | Electricity Mw |
|---|---|---|---|---|
| $SO_2$ Absorption | 25 | 77 | 25 | 10.1 |
| Solids Separation | - | - | - | 0.1 |
| Solids Disposal | - | - | 11 | 0.1 |
| TOTAL | 25 | 77 | 36 | 10.3 |

Source: (1,p.344)



TABLE 4.8

Raw Material and Energy Consumption for Lime Scrubbing

Basis: 500 Mw, 3.5 Percent Sulfur Coal, 5,6260 hr/yr

| Process Area | Lime tons/hr | Steam MM Btu/hr | Water M gal/hr | Electricity Mw |
|---|---|---|---|---|
| $SO_2$ Absorption | 11.6 | 77 | 25 | 9.5 |
| Solids Separation | - | - | - | 0.1 |
| Solids Disposal | - | - | 10 | 0.1 |
| TOTAL | 11.6 | 77 | 35 | 9.7 |

Source: (1,p.345)

TABLE 4.9

LIMESTONE SLURRY PROCESS
SUMMARY OF ESTIMATED FIXED INVESTMENT[a]

(500-MW new coal-fired power units, 3.5% S in coal; 1.2 lb $SO_2$/MBtu heat input allowable emission; onsite solids disposal)[a]

| Direct Investment | Investment, $ | Percent of total direct investment |
|---|---|---|
| Materials handling (hoppers, feeders, conveyors, elevators, bins, shakers, puller) | 1,759,000 | 6.8 |
| Feed preparation (feeders crushers, ball mills, hoists, tanks, agitators, and pumps) | 1,740,000 | 6.7 |
| Gas handling (common feed plenum and booster fans, gas ducts and dampers from plenum to absorber, exhaust gas ducts and dampers from absorber to reheater and stack) | 4,316,000 | 16.6 |
| Absorption (4 TCA scrubbers including humidification chambers and mist eliminators, effluent hold tanks, agitators, and pumps) | 8,916,000 | 34.3 |
| Stack gas reheat (4 indirect steam reheaters) | 1,262,000 | 4.9 |
| Solids disposal (onsite disposal facilities including feed tank, agitator, slurry disposal pumps, and pond water return pumps) | 1,656,000 | 6.4 |
| Subtotal | 19,675,000 | 75.7 |
| Services, utilities, and miscellaneous | 1,180,000 | 4.5 |
| Total process areas excluding pond construction | 20,855,000 | 80.2 |
| Pond construction | 5,145,000 | 19.8 |
| Total direct investment | 26,000,000 | 100.0 |
| Indirect Investment | | |
| Engineering design and supervision | 1,207,000 | 4.6 |
| Architect and engineering contractor | 260,000 | 1.0 |
| Construction expense | 3,617,000 | 13.9 |
| Contractor's fees | 1,142,000 | 4.4 |
| Total indirect investment | 6,234,000 | 23.9 |
| Contingency | 6,447,000 | 24.8 |
| Total fixed investment | 38,681,000 | 148.7 |
| Other Capital Charges | | |
| Allowance for startup and modifications | 3,354,000 | 12.9 |
| Interest during construction | 4,652,000 | 17.9 |
| Total depreciable investment | 46,677,000 | 179.5 |
| Land | 1,030,000 | 4.0 |
| Working capital | 1,021,000 | 3.9 |
| Total capital investment | 48,728,000 | 187.4 |

a. Basis:
Midwest plant location represents project beginning mid-1977, ending mid-1980. Average cost basis for scaling, mid-1979. Stack gas reheat to 175°F by indirect steam reheat. Minimum in-process storage; only pumps are spared. Disposal pond located 1 mi from power plant. Investment requirements for flyash removal and disposal excluded; FGD process investment estimate begins with common feed plenum downstream of the ESP. Construction labor shortages with accompanying overtime pay incentive not considered.

Source: (2,p.56)



TABLE 4.10

LIMESTONE SLURRY PROCESS
TOTAL AVERAGE ANNUAL REVENUE REQUIREMENTS-REGULATED UTILITY ECONOMICS[a]

(500-MW new coal-fired power unit, 3.5% S in coal; 1.2 lb $SO_2$/MBtu heat input allowable emission; on site solids disposal)[a]

| | Annual quantity | Unit cost, $ | Total annual cost, $ | Percent of total annual revenue requirements |
|---|---|---|---|---|
| Direct Costs | | | | |
| Raw materials<br>Limestone | 158,300 tons | 7.00/ton | 1,106,100 | 7.86 |
| Total Raw Material Cost | | | 1,106,100 | 7.86 |
| Conversion Costs<br>Operating labor and supervision | 25,990 man-hr | 12.50/man-hr | 324,900 | 2.30 |
| Utilities<br>Steam<br>Process water<br>Electricity | <br>489,300 MBtu<br>247,400 kgal<br>56,670,000 kWh | <br>2.00/MBtu<br>0.12/kgal<br>0.029/kWh | <br>978,600<br>29,700<br>1,643,400 | <br>6.94<br>0.21<br>11.66 |
| Maintenance<br>Labor and material | | | 1,822,600 | 12.93 |
| Analyses | 3,760 man-hr | 17.00/man-hr | 63,900 | 0.45 |
| Total Conversion Costs | | | 4,863,300 | 34.49 |
| Total Direct Costs | | | 5,971,400 | 42.35 |
| Indirect Costs | | | | |
| Capital Charges<br>Depreciation, interim replacements, and insurance at 6.0% of total depreciable investment | | | 2,800,600 | 19.86 |
| Average cost of capital and taxes at 8.6% of total capital investment | | | 4,190,600 | 29.72 |
| Overheads<br>Plant, 50% of conversion costs less utilities | | | 1,105,800 | 7.84 |
| Administrative, 10% of operating labor | | | 32,500 | 0.23 |
| Total Indirect Costs | | | 8,129,500 | 57.65 |
| Total Annual Revenue Requirements | | | 14,100,900 | 100.00 |
| | Mills/kWh/ | $/ton coal burned | $/MBtu heat input | $/short ton S removed |
| Equivalent unit revenue requirements | 4.03 | 9.40 | 0.45 | 408.01 |

a. Basis
Midwest plant location, 1980 revenue requirements.
Remaining life of power plant, 30 yr.
Power unit on-stream time, 7,000 hr/yr.
Coal burned, 1,500,100 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 175°F.
S removed, 34,560 short ton/yr; solids disposal 192,000 tons/yr calcium solids including only hydrate water.
Investment and revenue requirement for removal and disposal of flyash excluded.
Total direct investment, $26,000,000; total depreciable investment, $46,677,000; and total capital investment, $48,728,000.

Source: (2,p.56)

TABLE 4.11

Distribution of Estimated Captial Costs
for FGD Systems

| Category | Process | Limestone Wet Scrubbing 500 mw (1972) | Double Alkali Scrubbing 500 mw (1974) | Magnesia Scrubbing 500 mw (1972) | Sodium(W-L) Scrubbing 500 mw (1975) |
|---|---|---|---|---|---|
| I. | Gas handling area: fans, reheat system, scrubbers, duct work | 3,401,000 | 3,072,000 | 4,483,000 | 7,695,000 |
| II. | Chemical System | 293,000 | 2,808,000 | 2,734,000 | 6,840,000 |
| III. | Sulfuric acid plant | --- | --- | 2,877,000 | 2,565,000 |
| IV. | Bulk materials: Steel, concrete, piping, electrical & instrumentation work, paint, optional bypass duct, etc. | 2,777,000 | 2,521,000 | 3,411,000 | 7,600,000 |
| V. | Construction Costs: Labor, indirect field costs, construction fees | 5,513,000 | 5,800,000 | 3,796,000 | 8,740,000 |
| VI. | Contractor's Costs: Design engineering, fees, contingency, start-up, etc. | 3,800,000 | 3,300,000 | 6,361,000 | 4,560,000 |
| VII. | Interest during Construction | 629,000 | --- | 865,000 | --- |
| VIII. | Solids Disposal | 934,000 | --- | --- | --- |
| XI. | Yard Improvements | 375,000 | --- | 350,000 | --- |
| | Totals: | 17,722,000 | 17,501,000 | 24,877,000 | 38,000,000 |



TABLE 4.12

Distribution of Estimated Operating Costs
for FCD Systems

| Cost Category | Process | Limestone Wet Scrubbing 500 mw (1972) | Double Alkali Scrubbing 500 mw (1974) | Magnesia Scrubbing 500 mw (1972) | Sodium(W-L) Scrubbing 500 mw (1975) |
|---|---|---|---|---|---|
| | Direct Costs: | | | | |
| I. | Chemicals Conversion Costs | $ 394,600 | $2,193,378 | $ 137,000 | $ 956,000 |
| II. | Operating Labor & Supervision | 139,700 | 638,074 | 235,200 | 216,000 |
| III. | Utilities | 542,700 | 1,316,027 | 1,161,200 | 1,500,000 |
| IV. | Maintenance | 1,149,100 | 438,676 | 1,478,600 | 842,000 |
| V. | Laboratory (Analyses) | 38,000 | --- | 85,000 | 72,000 |
| VI. | Misc. Supplies | --- | --- | --- | 45,000 |
| | Subtotal Direct Costs: | 2,264,100 | 4,586,155 | 3,117,200 | 3,631,000 |
| | Indirect Costs: | | | | |
| VII. | Capital Charges | 2,140,200 | 1,236,268 | 3,721,500 | 2,712,000 |
| VIII. | Overhead | 342,400 | --- | 923,600 | 343,000 |
| | Subtotal Indirect Costs: | 2,482,600 | 1,236,268 | 4,645,300 | 3,055,000 |
| XI. | On Site Solids Disposal | 629,600 | Unknown | None | None |
| | Total Annual Operating Costs | $5,376,300 | $5,822,423 | $7,762,500 | $6,140,000 (no sulfur credit) |

Source: (3,p.56)

TABLE 4.13

Raw Material and Utility Requirements for the Wellman-Lord Process

Basis: 500 Mw, 3.5% Sulfur Coal, 5,260 Hours Per Year at Full Load

| Processing Area | Electric Power (Mw) | Steam-H.P. (MM Btu/hr) | Steam-L.P. (MM Btu/hr) | Reducing Gas (MM Btu/hr) | Sodium Carbonate (lb/hr) | Process Water (mgph) | Cooling Water (mgph) |
|---|---|---|---|---|---|---|---|
| Pretreatment and $SO_2$ Absorption | 9.3 | 77 | - | - | 2,120 | - | - |
| Regeneration | 1.6 | - | 160 | - | - | 25.5 | 430 |
| Purge Treatment | 1.0 | - | 20 | - | - | - | - |
| $SO_2$ Conversion | 0.3 | - | -(12) | 92 | - | - | - |
| TOTAL | 12.2 | 77 | 168 | 92 | 2,120 | 25.5 | 430 |

Source: (1,p.256)



TABLE 4.14

Comparative Capital Investments for Power Systems ($MM)

| | PC Boiler with Flue Gas Desulfurization | Atmospheric FBC Boiler | Pressurized FBC Boiler |
|---|---|---|---|
| Equipment Costs | | | |
| Steam gen./combustor | 31.5 | 25.1 | 14.5 |
| $SO_2/NO_x$ control | 30.0 | 3.7 | 14.5 |
| Particulate removal | 10.0 | 10.0 | 12.6 |
| Coal handling & feeding | 13.6 | 11.0 | 10.5 |
| Piping, flue, & ducts | 2.6 | 1.7 | 4.4 |
| Air heater | 3.3 | 4.6 | 0.0 |
| Other boiler plant | 7.6 | 9.7 | 6.0 |
| Total Boiler Plant/Combustor | 98.6 | 65.6 | 52.7 |
| Gas Turbine-Generator | - | - | 14.6 |
| Steam Turbine-Generator | 55.8 | 55.0 | 44.1 |
| Structures & Improvements | 27.3 | 24.8 | 18.5 |
| Other Plant Consts.(electrical) | 53.2 | 53.3 | 49.3 |
| Total Equipment | 234.1 | 196.9 | 179.4 |
| Contingency | 14.1 | 11.9 | 10.8 |
| Interest & Escalation | 107.2 | 86.4 | 65.0 |
| General & Engineering | 5.5 | 5.5 | 5.5 |
| Total | 352.9 | 302.7 | 260.7 |

NOTE - 500 MW plant, 1976 dollars, throwaway products from FGD and FBC

Source: (1,p.5-13)

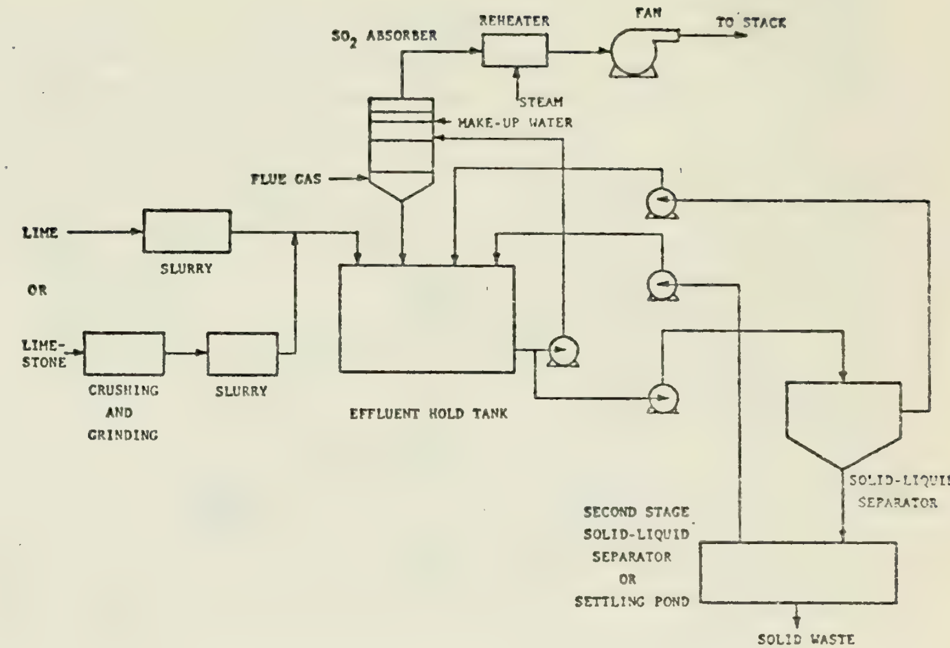
| TABLE 4.15<br>Comparative Operating Costs for Power Systems ($MM/KwH) | | | |
|---|---|---|---|
| | PC Boiler with Flue Gas Desulfurization | Atmospheric FBC Boiler | Pressurized FBC Boiler |
| Capital Charges @ 15 percent | 7.56 | 6.48 | 5.36 |
| Maintenance & Operating | 0.99 | 0.67 | 0.71 |
| Fuel @ 40¢/MMBTU | 3.60 | 3.60 | 3.60 |
| Dolomite or Limestone (makeup & disposal) | 0.60 | 1.10 | 1.10 |
| TOTAL | 12.75 | 11.82 | 10.77 |

NOTE - 7000 hrs/year

Source: (1,p.5-14)



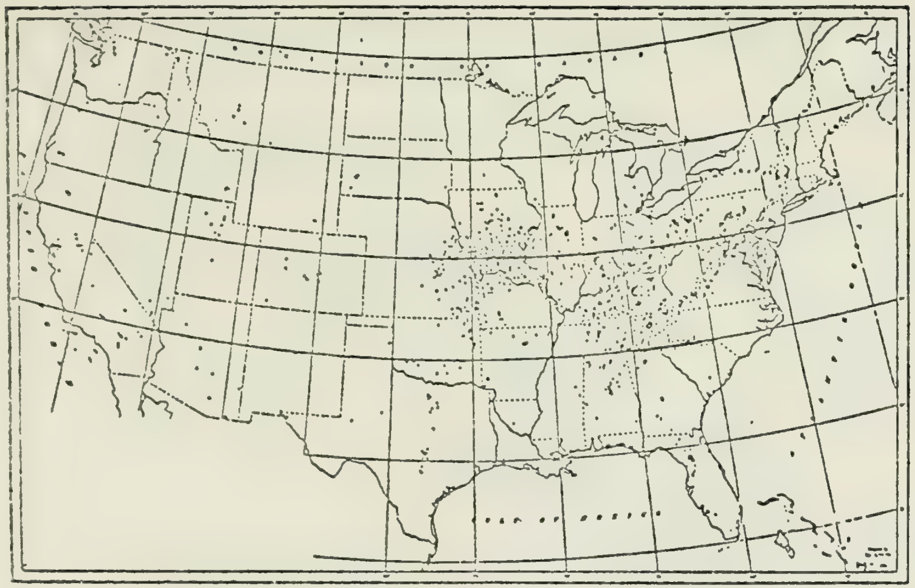
FIGURE 4.1.




Process Flow Diagram for Lime/Limestone Scrubbing Process

Source: (1,p.337)

FIGURE 4.2

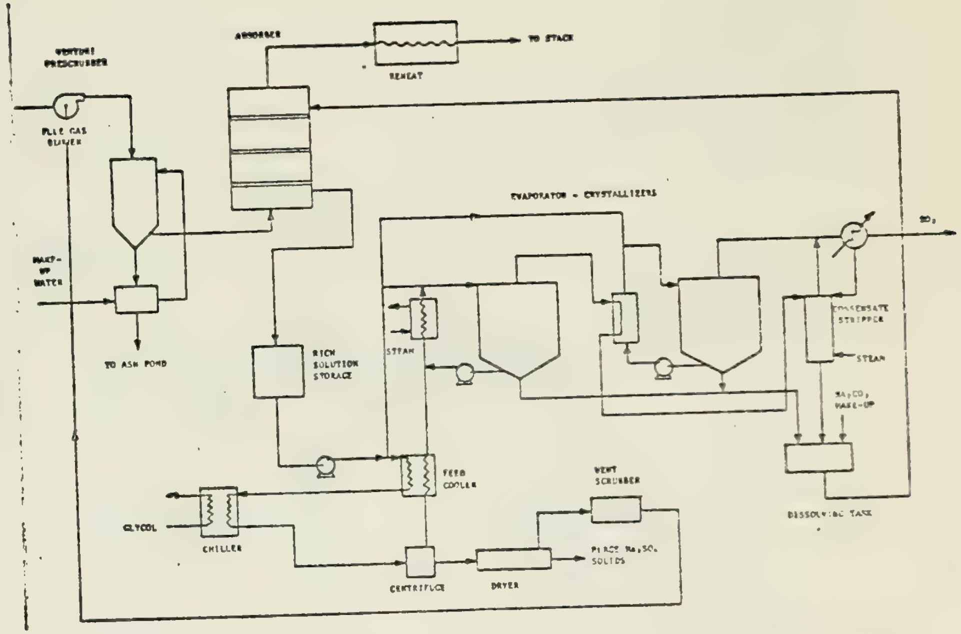


Location of Active High-Calcium Limestone Deposits in the United States

Source: (2,p.36)



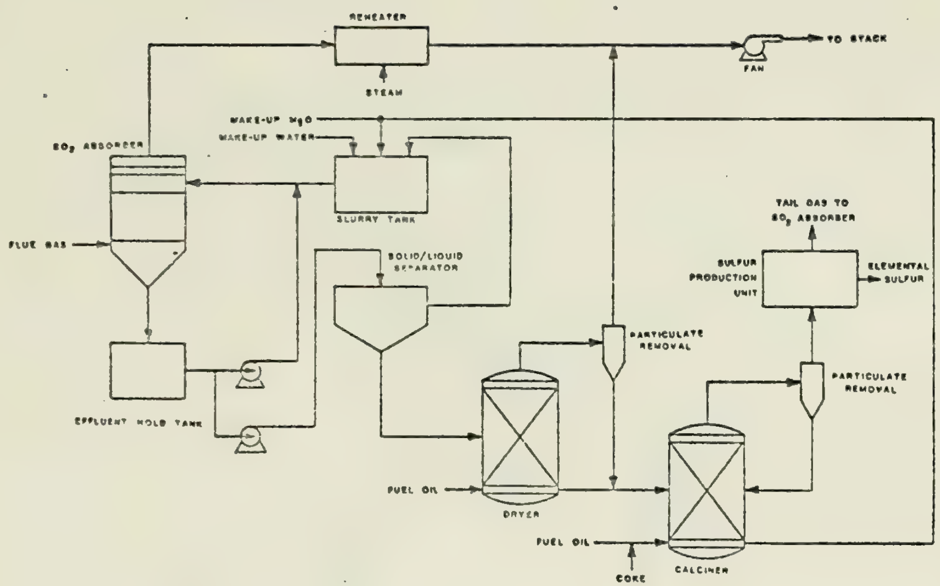
FIGURE 4.3.




Wellman-Lord Process Flow Diagram

Source: (1,p.247)

FIGURE 4.4.

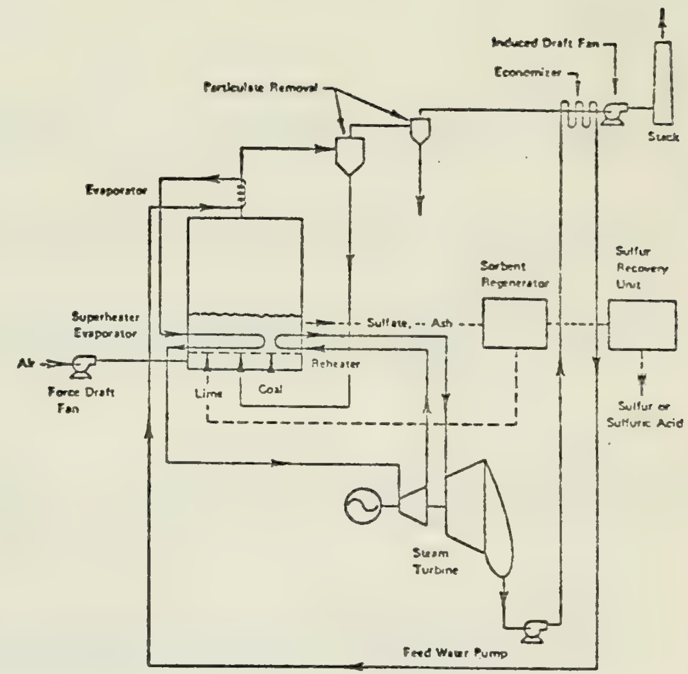



Process Flow Diagram for the Magnesia Slurry Absorption Process

Source: (1,p.311)



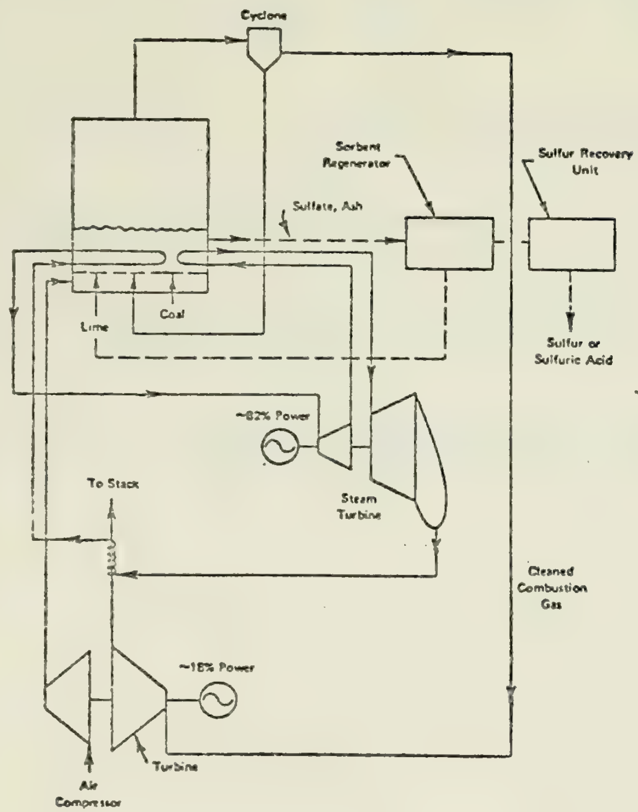
FIGURE 4.5




Atmospheric Fluidized-Bed Combustion Power Plant

Source: (1,p.5-7)

FIGURE 4.6




Pressurized Fluidized-Bed Combustion Power Plant

(1,p.5-8)



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# 5.0 EXTRA HIGH VOLTAGE TRANSMISSION AND DISTRIBUTION

There are three possible modes for moving coal-energy to and from the ORBES region. They are rail transport, slurry pipelines, and extra high voltage transmission. The first two methods deal with the physical movement of coal, whereas the third deals with moving the power product. In extra high voltage (EHV) transmission the coal is transformed into electricity in the mining region and sent to distant consumers. The feasability of this idea was demonstrated during the coal strike this past winter.

There are two types of power movement available - AC and DC transmission. AC transmission has been our leading power mover since the development of the AC transformer. This allowed the transport of large voltages over greater distances. But there are characteristic line losses in this system which can be as much as 33 percent greater than for a comparable DC system [1,p.90]. Improved insulator and conductor bundle design may improve the transmission efficiency of an AC system but, at this time, it appears that for distances greater than 600 miles the DC system would be a better choice (even with the high cost of terminal equipment). This is illustrated in Figure 5.1 [1,p.95]. Some of the more desirable features of a DC system are given below [1,p.91]:

1. Greater power per conductor

2. Lighter and simpler line and tower construction

3. A ground return can be used-each conductor can be operated as an independent circuit

4. No charging current

5. No skin effect

6. The cables can be worked at a higher voltage gradient

7. The line does not require reactive compensation

8. There is less corona loss and interference

9. Distance is not limited by stability

10. A DC system may interconnect with AC systems of different frequencies

11. There is low short circuit current

12. Tie-line power is easily controlled

It appears that a DC transmission system would be the obvious choice for delivering bulk power loads over great distances. Tables 5.1 (1,p.98) and 5.22 (1,p 99) list some relative costs for representative AC and DC transmission systems, respectively. Note that the capital investments for a DC transmission system are substantially lower, as are the annual fixed costs. Although



the operating costs are slightly higher, the overall annual charges do indicate a rather large potential savings for this distance.

For the shorter distances-such as distribution lines within the ORBES region - AC systems would be preferable. The primary difference between transmission and distribution systems is simply their function [2,p.I-1]. Transmission systems deal with moving large blocks of power to major load centers, whereas distribution systems are responsible for voltage reduction and circuit subdivision. Thus, large demands could effectively be met by an AC grid, but the relatively lower voltages would reduce their inherent, and uneconomical, line losses.

A DC transmission system feeding into an AC distribution network would appear to be the optimum design. But the key concept here is that of decentralization of the power supply. This system not only has the ability to import to, or export from, the ORBES region large amounts of power with minimal line loss, but it can also direct this voltage in substantial amounts through numerous regional interconnects. This provides greater mobility, at the lowest cost, for supplying various amounts of power to different sectors of a region. To go a step further, future power plant reserve capacities could be looked at as an overall network supply and could be reduced.

TABLE 5.1

COST ANALISIS OF 765 KV* EHV AC TRANSMISSION SYSTEM
(1975 dollars)
(3000 MW, 1000 Mile)

| | | | | |
|---|---|---|---|---|
| I. | CAPITAL INVESTMENTS: | | | |
| | A. Sending Substation | 35,651,600 | | |
| | B. Receiving Substation | 41,306,900 | | |
| | C. Compensation | | | |
| | C1. Series Capacitors @ $14.0/kvar | 59,500,000 | | |
| | C2. Shunt Reactors @ $18.6/kvar | 79,050,000 | | |
| | C3. Shunt Capacitors @ $10 65/kvar | 15,442,500 | | |
| | D. Transmission Line Facilities @ $353,690/mile | 353,690,000 | | |
| | E. Right of Way @ $1500/acre | 40,500,000 | | |
| | Total Capital Costs: | | 625,141,000 | |
| II. | ANNUAL FIXED COSTS:* | | | |
| | A. Sending Substation @ 13.9% of IA | 4,955,572 | | |
| | B. Receiving Substation @ 13.9% of IB | 5,741,659 | | |
| | C. Compensation @ 13.9% of IC | 21,404,961 | | |
| | D. Transmission Facilities @ 13.5% of ID | 47,748,167 | | |
| | Total Annual Fixed Costs: | | 79,850,339 | |
| III. | ANNUAL OPERATING COSTS:** | | | |
| | A. Sending Substation @ 2.34% of IA | 834,247 | | |
| | B. Receiving Substation @ 2.60% of IB | 1,073,979 | | |
| | C. Compensation @ 1.3% of IC | 2,001,902 | | |
| | D. Transmission Facilities @ 1.3% of ID | 4,597,968 | | |
| | E. Electrical Energy Losses: | 37,928 | | |
| | E1. Energy Losses @ 1¢/kwh | | | |
| | E2. Power Losses @ $180/kw & 13.9% fixed chgs | | | |
| | Total Annual Operating Costs: | 8,546,024 | | |
| IV. | ANNUAL COSTS: (II + III) | | 88,396,363 | |
| V. | UNIT COSTS: | | | |
| | A. Mills/ton-mile of equivalent coal | | | 7.641 |
| | B. Mills/kw-hr received power | | | 3.843 |
| | C. Investment $/kw-hr received power | | | 27.2 |

*Double circuit is used.

*Based on FPC P-38 Annual fixed charge rate.

**Includes administrative and general expenses @ 3.0% of their operating costs.

Source: (1,p.98)



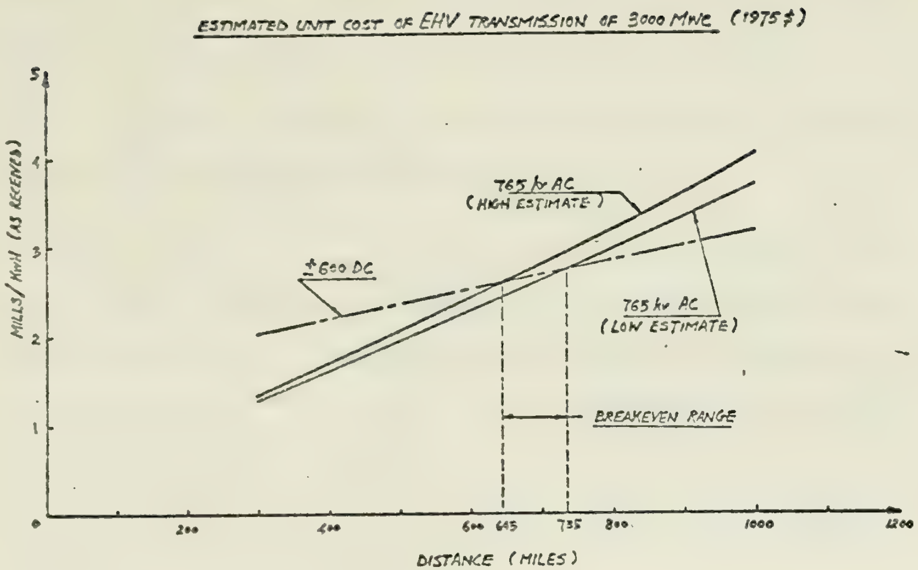
TABLE 5.2

COST ANALYSIS OF 765 KV EHV-DC TRANSMISSION SYSTEM
(1975 dollars)
(3000 MW, 1000 Mile)

| | | | | |
|---|---|---|---|---|
| I. | CAPITAL INVESTMENTS: | | | |
| | A. Sending Substation | 123,000,000 | | |
| | B. Receiving Substation | 118,712,000 | | |
| | C. Transmission Line Facilities @ $236,000/mile | 236,000,000 | | |
| | D. Right of Way @ $1500/acre | 31,500,000 | | |
| | Total Capital Costs: | | 509,212,000 | |
| II. | ANNUAL FIXED COSTS:* | | | |
| | A. Sending Substation @ 13.9% of IA | 17,097,000 | | |
| | B. Receiving Substation @ 13.9% of IB | 16,501,000 | | |
| | C. Transmission Facilities @ 13.5% of ID | 31,860,000 | | |
| | Total Annual Fixed Costs: | | 65,458,000 | |
| III. | ANNUAL OPERATING COSTS:** | | | |
| | A. Sending Substation @ 2.34% of IA | 2,878,200 | | |
| | B. Receiving Substation @ 2.60% of IB | 3,086,500 | | |
| | C. Transmission Facilities @ 1.3% of ID | 3,068,000 | | |
| | D. Electrical Energy Losses: | 30,600 | | |
| | D1. Energy Losses @ 1¢/kwh | | | |
| | D2. Power Losses @ $80/kw & 13.9% fixed chgs | | | |
| | Total Annual Operating Costs: | | 9,063,300 | |
| IV. | ANNUAL COSTS: (II + III) | | | 74,521,300 |
| V. | UNIT COSTS: | | | |
| | A. Mills/ton-mile of equivalent coal | | | 6.269 |
| | B. Mills/kw-hr received power | | | 3.153 |
| | C. Investment $/kw-hr received power | | | 21.5 |

Source: (1,p.99)

FIGURE 5.1




Estimated Unit Cost of EHV Transmission of 3000 MWe (1975$)

Source: (1,p.95)



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## 6.0 RECOMMENDATIONS - I-O ANALYSIS

The methodology applied here is that of Herendeen (1) with handbook examples provided by Bullard (2). This involves constructing column matrices of technical coefficients showing the relative operating costs of future coal technologies. As these are energy coefficients, they are expressed in Btu/Btu for energy sectors and $/Btu for non-energy sectors.

The first step is the categorization within the Ohio River Basin Energy Study region (ORBES) of the costs involved in the various technologies. The ORBES demand study has established a 48 by 48 coefficient matrix which separates and aggregates various aspects of the ORBES region economy. The SIC classifications for the ORBES IO model are shown in Table 6.1. From this list it is determined which classifications apply to the energy technologies considered here.

The next step is to examine the subdivisions of the applicable SIC classifications to determine the cost considerations more specifically. Up to this point all eliminations have been due simply to their inapplicability to any of the processes considered. That is, nothing has been discounted due to specific selection of costs requiring documentation. Note also that in Table 6.1 the ORBES



classifications are listed such that the SIC classifications are divided into three groups listed in numerical order. Each group exhibits a decreasing energy sensitivity in relation to the ORBES economy. Thus no SIC classifications were subdivided past ORBES group number 30; costs past this point, where applicable, are relatively constant. In addition, the second sequence contains many stable costs. This is especially true of groupings numbered 5000 and above (wholesale and retail trade, finance, insurance, real estate and services). Therefore, the major differences among the various technologies occurs in ORBES groups 1-5, which include SIC classifications 1100 to 5000 (coal mining, construction, manufacturing, transportation, communication, and electricity, gas, and sanitary services).

The evaluation procedure to be followed for the relatively insensitive categories is analagous, where more specific disaggregation is utilized in distinguishing the new technologies. For example, consider furniture requirements for the offices associated with plants. Because commercial plants, e.g. for gasification or liquefaction, are still in the future, specific furniture requirements can only be approximated. For this insensitive category, sufficient approximation can be obtained by utilizing the office furniture requirements for a similar type of plant operation. This procedure can also be applied to categories such as heating, some construction,

transportation, etc.

Alternatively, consider low-Btu gasification processes utilizing a fluidized-bed, air-blown, ash-agglomerating process as the most feasible future technique. Currently the major drawback with these processes are operating characteristics involving coal feed and fines handling which can affect the efficient use of the coal. Thus specific consideration must be given to related materials and labor requirements. However, the use of low-Btu processes will be in large industrial parks for process heat, or as a boiler fuel in steam power plants. Thus many of the costs are analagous to contemporary steam generator operation and supervision. In addition, these costs may be extrapolated to high-Btu gasification operations (also utilizing a fluidized bed) to include oxygen requirements, shift conversion, and methanation steps. The most common plant size studied is one producing 250 MMscfd of gas.

A similar analogy can be drawn for liquefaction and other processes. In the case of liquefaction a chemical processing plant can be used, or a refinery producing 50,000 bpd. The use of a refinery is advantageous in that a potential design application of liquefaction is the "coalfinery" described previously. Fluidizesd bed combustion also utilizes steam generation facility data, with coal handling and dolomite supply as additional operational costs. Combined cycle configurations



simply utilize gasification cost results and incorporate them with given data on steam and gas turbine power plants. The same may be said for general applications of cogeneration units. For lime/limestone FGD systems, some adjustments, such as $M_gO$ supply, have to be made for the magnesia slurry technique. Extra high voltage transmission costs can be approximated from existing data, as essentially conventional technology used for AC or DC systems.

Upon selecting the technologies with potential commerciality, the problem becomes one of determining the costs incurred in these processes. This involves an economic and engineering analysis of the necessary equipment. More importantly, it involves discriminatory utilization of applicable verified information. New ideas are still founded on what is currently available. The next step is separation by process of the general costs (for example, coal mining would not be a cost incurred by an FGD system). From this energy costs may be determined in the appropriate manner for sensitive and insensitive categories. This analysis is facilitated by a Brookhaven report by Lukachinski and Tessmer (3). This report contains technical coefficients for energy technologies projected to 1985. This is appropriate for our purposes as 1985 may be taken as a key transition point for the processes proposed. The non-energy coefficients are normalized in terms of 1967 dollars. The basic IO energy study utilized a 110-sector economy as shown

in Table 6.2 (3,p.3). However, this in turn was reduced from a 367-sector IO energy matrix based on BEA data. The 48-sector ORBES energgy demand study is consistent with the above. This makes aggregation a simple case of bookkeeping. Much of the information given in the report is based on the aforementioned MITRE study (4), which is also subdivided according to the 367-sector BEA economy. In some cases this study are referred to directly. This chapter only discusses sectors undergoing change. Other sectors' documentation will simply be referenced, as they are included in the overall ORBES matrix.

As an example, consider high-Btu coal gasification. The method used in the Brookhaven report was the Hygas process, which is the representative process chosen in this report for the future. The coefficients given in this report (3,p.18) remain essentially the same for our purposes, especially the non-energy requirements. One possible change could be that of the chemical feedstock but, as shown in Table 2.9, this makes up less than 3% of the cost and may be taken as given. The change that must be made is in the type of coal used. We are interested in the bituminous coal requirements rather than those of lignite. This can be demonstrated by proportionately adjusting the Btu requirement of input coal in terms of relative Btu ratings of equal tonnage of bituminous and lignite coal. Comparing the dry heating values of Illinois bituminous and lignite coal (5,p.13), it is seen that the coal requirement should be reduced by 16%.



It is conceded that lignite is more reactive and cleaner, but it must be assumed that operational problems are resolved once the process is ready for commercial application.

The representative low-Btu gasification process being considered is U-Gas. Both Hygas and U-Gas are techniques developed by IGT and have many similar operational and design characteristics, resulting in a good deal of similarity in the technical coefficients. The Hygas coefficients are used as a baseline for this analysis. For example, coal handling and pretreatment costs are similar for both processes, as well as purification costs. At first one might think coal requirements would be drastically different. This, however, is not the case as both Table 1.3 and the MITRE report (4,p.220) show similar operational efficiencies. Naturally U-Gas requires more coal recycling to obtain as much Btu output, but this again is part of the resolvable operational considerations. The important differences in coefficients occur due to hydrogen production and methanation requirements. Because CO conversion and methanation catalysts are removed for the low-Btu gasification process, they may be eliminated from the chemical requirements (4,p.210) and hence reduce section 50 of the Brookhaven model by 54%. Note that the adjustment has again been made proportionately, as we are dealing with ratios. The hydrogen-producing technique used in the given study was the electrothermal method. Although this may not be the technique used in the future, elimination of this

can give a representative reduction in cost. Elimination of the Btu requirement for power generation (assuming the majority is required for the electrothermal process) results in a coal coefficient (Sector 1) reduction of 6%. Note that this is applied to the coal sector, as coal is used for the Btu input. Recall that this modification is applied to the changed Hygas coefficient mentioned above.

Brookhaven also conducted a study of the SRC liquefaction process (3,p.20), utilizing much of the MITRE data. Here the SRC product has been considered as a boiler fuel product, which is where much of the emphasis is being placed for the proposed demonstration plants. As mentioned before, the product may later be used as a refinery feedstock, which may require some of the coefficients to be altered and/or incorporated into a refinery operation. The non-energy coefficients given are representative for our purposes at this time. In addition, because a Pittsburgh seam coal is used for this study, and refinery and coal gasification coefficients are used where applicable, the energy coefficients are also useful for our purposes.

But this only demonstrates the requirements for a solvent extraction technique (material is not yet available, but it may be assumed that the coefficients for the Exxon Donor-Solvent technique are similar to those of SRC). For a hydrogenation technique, such as H-coal, a few changes in the coefficients are



required. As the principle applied is generally the same for both techniques, we will assume similar non-energy coefficients. The major differences are due to larger coal (Sector 1) and chemical (Sector 50) requirements. This is demonstrated by the requirements given in Tables 2.20 and 2.23. Assuming the same cost and characteristics for the materials required, these two sectors may be adjusted proportionately. The coefficient for coal increases 15% and the sector for chemicals by 159% over those of SRC. All other coefficients are considered applicable as given.

The technical coefficients for a limestone slurry process, the most extensively used FGD system, are taken from Table 4.10. The lime slurry FGD system is also used, but its requirements are essentially the same as those for limestone. More operational information is required to do a study of the regenerative magnesia slurry technique, but Table 4.11 illustrates some of the relative cost considerations. Note the savings for solids disposal.

It must be pointed out that the FGD system is not an entity by itself. Its coefficients are incorporated into the operation of a plant where cleanup is required. The coefficients for a limestone process, derived from data in Table 4.10 in terms of the 110-sector model, are given below:

| | | |
|---|---|---|
| (16) | .0409 | Btu/Btu |
| (20) | .0162 | Btu/Btu |
| (35) | .2803 | $/MMBtu |
| (50) | .0927 | $/MMBtu |
| (100) | .5851 | $/MMBtu |

The technical coefficients presented by the Brookhaven model for combined cycle operation seem to be the best available. These are incorporated from MITRE data as explained in their report. This involves the incorporation of a COGAS topping system with an entrained bed low-Btu gasification system. This topping system is representative of possible future application. In addition, a hot-gas cleanup system is utilized, which must be assumed economically feasible as high temperature gases are required if substantial output is to be available. Note that we assume resolution of the high-temperature turbine blade problems. A fluidized-bed gasifier may be preferred to the entrained-bed variety, but many of the advantages are present with this system and are reflected in the coefficients. For comparison a coefficient from Sector 1 was determined for the EAH case of Table 3.8 for total potential output (i.e., 1000 Mw) and was found to be 2.108.

In the Brookhaven report (3,p.16), the technical coefficients for an FBC system are essentially the same as those for a coal-fired electric generation plant. Note that these



coefficients are for utility boiler use. The major difference here occurs for maintenance and operation (Sector 35) and chemical requirements (Sector 50). These adjustments are made from relative costs given in Table 4.15. This table, taken from a Teknekron paper (6.p.5-14), compares a pulverized bed combustion boiler to atmospheric and pressurized fluidized bed systems. Fuel costs have been normalized to equity in that table; that is, efficiency advantages are incorporated with maintenance and operating considerations. Thus there is no change in Sector 1; we assume the same Btu values are extracted in each case at the expense of increased operating costs. It must be remembered that the values given in this table are relative values, but the initial coefficients from the Brookhaven report are viewed as reasonably accurate. From Table 4.15 it is seen that Sector 35 will decrease by 32% for the atmospheric case and 28% for the pressurized case. Sector 50 will increase by 83% for both cases. Note that it is assumed here that the greatest effect on changing technologies is operational procedure. Note also that an additional change will occur under finance and insurance charges (Sector 100). This will decrease by 14% for the atmospheric case and 29% for the pressurized case.

EHV transmission may be considered here as a cost of utilization or a mode of potentially competitive imports or exports. That is, aside from fixed charge rates on the equipment, and manual operational expenses, there is no

technological breakdown necessary for annual operating costs. This is simply a process for delivering generated power from one location to another, and is entered in toto under Sector 77 of the Brookhaven model. We use as a standard example (the values given by Soo and Rieber (7,p.99) in Table 5.2) for a ±600KV EHV-DC transmission system delivering 3000 MW over 1000 miles. The values in this table result in a technical coefficient with a value of 1.03944 dollar per million Btus.

The values used here are the result of rather extensive and detailed studies of applicable coal technologies. The changes made not only alter these coefficients for our specific purposes, but also illustrate some of the more important considerations related to the operating costs of these processes. Thus we not only selected a generic process with feasible future applications from among the prominent technologies available, but have been able to give a comparable prediction of costs for these technologies. Recall again that we assumed resolution of various equipment and operational problems in some cases. For example, we assumed development of an effective hot-gas cleanup system and a high-temperature turbine blade for a combined cycle power plant. In addition we assumed development of a pressurized fluidized bed reactor, with favorable results compared to the atmospheric case. The main argument against the processes in this study is the actual time horizon we are considering. This also applies to the future regenerable FGD systems as much of the



future plans involve contemporary "throw-away" techniques. Hopefully this trend may change and make direct combustion systems attractive and economically feasible. This is especially critical in the high-demand, high-sulfur coal, ORBES region.

But this very demand may dictate development and advocacy of coal conversion technologies. High-Btu gasification, the original conversion technology on which extensive research and development had been undertaken, appears to be the most expensive for what is received. It can only be justified on an extensive scale by a desperate demand for natural gas alone. Low-Btu gasification is much more economical and could be of great benefit in industrial as well as utility applications. A major possiblity is the incorporation of a process with an industrial park. Note that the most favorable processes for gasification are of the fluidized-bed variety, as these are able to handle the predominately caking coals of the ORBES region.

Another possbility for a gasification system is its incorporation into a coal liquefaction complex. For example, it could extract additional coal products from the residual char, as was the original purpose of the COGAS technique. The most desirable future coal conversion technology for the ORBES region appears to be coal liquefaction; specifically the SRC (solvent extraction) technique. This seems to be the most economical process per unit of output. It has a high conversion efficiency,

and can handle the high-sulfur, caking coals of the ORBES region. It is also a proven process soon to be demonstrated on a large scale. Currently these plants will be designed to produce a clean boiler fuel, but future expansion should include the aforementioned multiproduct "coalfinery" concept. This would allow the ORBES region to deal not only with increases in energy demand, but variations in the fuels required in the future as well.

It is rather obvious that the technologies, in various stages of development, are present to assist in meeting the energy demands of the ORBES region. It is also obvious that development and the use of the extensive coal resources in the ORBES region can play an important part in the national energy economy. But it must be kept in mind that the type of energy demanded (e.g., for residential, industrial or utility purposes) will help dictate not only the process but the fuel required. This makes the coal liquefaction concept so attractive.



TABLE 6.1

SIC CLASSIFICATIONS FOR
48-SECTOR ORBES MODELS

| ORBES REGION | SIC CLASSIFICATIONS | | | | |
|---|---|---|---|---|---|
| 1 | 1100 | 1200 | | | |
| 2 | 1310 | 1320 | | | |
| 3 | 2910 | 2990 | | | |
| 4 | 4910 | 4931 | | | |
| 5 | 4920 | 4932 | | | |
| 6 | 01-- | | | | |
| 7 | 1000 | 1400 | | | |
| 8 | 15-- | 1380 | | | |
| 9 | 2010 | | | | |
| 10 | 2000 | -2010 | 2100 | | |
| 11 | 2200 | | | | |
| 12 | 2300 | | | | |
| 13 | 2400 | 2500 | | | |
| 14 | 2600 | -2620 | -2630 | -2650 | 2700 |
| 15 | 2620 | | | | |
| 16 | 2630 | | | | |
| 17 | 2650 | | | | |
| 18 | 2810 | | | | |
| 19 | 2800 | -2810 | -2820 | | |
| 20 | 2820 | | | | |
| 21 | 2950 | | | | |
| 22 | 3000 | 3100 | | | |
| 23 | 3200 | | | | |
| 24 | 3310 | | | | |
| 25 | 3320 | 3390 | | | |
| 26 | 3330 | | | | |
| 27 | 3340 | 3350 | 3360 | | |
| 28 | 3400 | | | | |
| 29 | 3500 | | | | |

TABLE 6.1

(Continued)

SIC CLASSIFICATIONS FOR 48-SECTOR ORBES MODELS

| ORBES REGION | SIC CLASSIFICATIONS | | | | |
|---|---|---|---|---|---|
| 30 | 3600 | | | | |
| 31 | 3710 | | | | |
| 32 | 3700 | -3710 | | | |
| 33 | 3800 | 3900 | | | |
| 34 | 4740 | | | | |
| 35 | 4100<br>4940 | 4600<br>4950 | 4700<br>4960 | -4740<br>4970 | 4800 |
| 36 | 4200 | | | | |
| 37 | 4400 | | | | |
| 38 | 4500 | | | | |
| 39 | 5000 | | | | |
| 40 | 52-- | 7396 | | | |
| 41 | 6000 | 6100 | 6200 | 6400 | 6700 |
| 42 | 6500 | 6600 | | | |
| 43 | 7000<br>7800 | 8100<br>7900 | 8900 | -8920 | |
| 44 | 7200 | 7600 | 7300 | -7310 | -7396 |
| 45 | 7310 | | | | |
| 46 | 7500 | | | | |
| 47 | 8000 | 8200 | | | |
| 48 | 8400 | 8600 | 8920 | | |



TABLE 6.2

SECTORAL CLASSIFICATION OF ENERGY INPUT-OUTPUT MODEL AND ALIGNMENT WITH OTHER CLASSIFICATION SYSTEMS

| BNL 110 | Sector | BNL 101 | BEA |
|---|---|---|---|
| 1 | Coal | 1 | 7 |
| 2 | Crude Oil & Gas | 2 | 8 |
| 3 | Shale Oil | - | - |
| 4 | Methane from Coal | 3 | - |
| 5 | Solvent Refined Coal | - | - |
| 6 | Refined Oil Products | 4 | 31.01 |
| 7 | Pipeline Gas | 5 | 68.02 |
| 8 | Coal Combined Cycle Electric | - | - |
| 9 | Other Fossil Electric | 6 | 68.01 |
| 10 | LWR Electric | 7 | 68.01 |
| 11 | HTGR Electric | - | - |
| 12 | Hydroelectric | 8 | 68.01 |
| 13 | Ore Reductions Feedstocks | 9 | - |
| 14 | Chemical Feedstocks | 10 | - |
| 15 | Motive Power | 11 | - |
| 16 | Process Heat | 12 | - |
| 17 | Water Heat | 13 | - |
| 18 | Space Heat | 14 | - |
| 19 | Air Conditioning | 15 | - |
| 20 | Electric Power | 16 | - |
| 21 | Livestock and Livestock Products | 17 | 1 |
| 22 | Other Agricultural Products | 18 | 2 |
| 23 | Forestry and Fishery Products | 19 | 3 |
| 24 | Agricultural, Forestry and Fishery Services | 20 | 4 |
| 25 | Iron and Ferroalloys Ores Mining | 21 | 5 |
| 26 | Nonferrous Metal Ores Mining | 22 | 6 |
| 27 | Stone and Clay Mining and Quarrying | 23 | 9 |
| 28 | Chemicals and Fertilizer Mineral Mining | 24 | 10 |
| 29 | New Construction, Residential Buildings | -- | 11.01 |
| 30 | New Construction, Nonresidential Buildings | -- | 11.02 |
| 31 | New Construction, Public Utilities | 25 | 11.03 |
| 32 | New Construction, Highways | -- | 11.04 |
| 33 | New Construction, All Other | -- | 11.05 |

Code BNL 110 - 110 sector version
BNL 101 - 101 sector version
BEA - Bureau of Economic Analysis, U.S. Dept. of Commerce

Source: (3,p.3)

TABLE 6.2
(Continued)

SECTORAL CLASSIFICATION OF ENERGY INPUT-OUTPUT MODEL AND ALIGNMENT WITH OTHER CLASSIFICATION SYSTEMS

| BNL 110 | Sector | BNL 101 | BEA |
|---|---|---|---|
| 34 | Maintenance and Repair Construction, Residential | — | 12.01 |
| | | 26 | -- |
| 35 | Maintenance and Repair Construction, All Other | — | 12.02 |
| 36 | Ordinance and Accessories | 27 | 11 |
| 37 | Food and Kindred Products | 28 | 14 |
| 38 | Tobacco Manufactures | 29 | 15 |
| 39 | Broad and Narrow Fabrics, Yarn and Thread Mills | 30 | 16 |
| 40 | Misc. Textile Goods and Floor Coverings | 31 | 17 |
| 41 | Apparel | 22 | 18 |
| 42 | Misc. Fabricated Textile Products | 33 | 19 |
| 43 | Lumber and Wood Products, Except Containers | 34 | 20 |
| 44 | Wooden Containers | 34 | 20 |
| 45 | Household Furniture | 36 | 22 |
| 46 | Other Furniture and Fixtures | 37 | 23 |
| 47 | Paper and Allied Products Except Containers and Boxes | 38 | 24 |
| 49 | Printing and Publishing | 40 | 26 |
| 50 | Chemicals and Selected Chemical Products | 41 | 27 |
| 51 | Plastics and Synthetic Materials | 42 | 28 |
| 52 | Drugs, Cleaning and Toilet Preparations | 43 | 29 |
| 53 | Paints and Allied Products | 44 | 30 |
| 54 | Paving Mixtures and Blocks | 45 | 31.02 |
| 55 | Asphalt Felts and Coatings | 46 | 31.03 |
| 56 | Rubber and Miscellaneous Plastics Products | 47 | 32 |
| 57 | Leather Tanning and Industrial Leather Products | 48 | 33 |
| 58 | Footwear and Other Leather Products | 48 | 33 |
| 59 | Glass and Glass Products | 49 | 34 |
| 60 | Stone and Clay Products | 51 | 36 |

Code BNL 110 - 110 sector version
BNL 101 - 101 sector version
BEA - Bureau of Economic Analysis, U.S. Dept. of Commerce
Source: (3,p.3)



TABLE 6.2
(Continued)

SECTORAL CLASSIFICATION OF ENERGY INPUT-OUTPUT MODEL AND ALIGNMENT WITH OTHER CLASSIFICATION SYSTEMS

| BNL 110 | Sector | BNL 101 | BEA |
|---|---|---|---|
| 61 | Primary Iron and Steel Manufacturing | 52 | 37 |
| 62 | Primary Nonferrous Metals Manufacturing | 52 | 38 |
| 63 | Metal Containers | 54 | 39 |
| 64 | Heating, Plumbing and Fabricated Structural Metal Products | 55 | 40 |
| 65 | Screw Machine Products, Bolts, Nuts, etc. & Metal Stampings | 56 | 41 |
| 66 | Other Fabricated Metal Products | 57 | 42 |
| 67 | Engines and Turbines | 58 | 43 |
| 68 | Farm Machinery | 59 | 44 |
| 69 | Construction, Mining, Oil Field Machinery, Equipment | 60 | 45 |
| 70 | Materials Handling Machinery and Equipment | 61 | 46 |
| 71 | Metalworking Machinery and Equipment | 62 | 47 |
| 72 | Special Industry Machinery and Equipment | 63 | 48 |
| 73 | General Industrial Machinery and Equipment | 64 | 49 |
| 74 | Machine Shop Products | 65 | 50 |
| 75 | Office, Computing and Accounting Machine | 66 | 51 |
| 76 | Service Industry Machines | 67 | 52 |
| 77 | Electric Trans. & Dist. Eq. & Electric Industry Apparatus | 68 | 53 |
| 78 | Household Appliances | 69 | 54 |
| 79 | Electric Lighting and Wiring Equipment | 70 | 55 |
| 80 | Radio, Television and Communications Equipment | 71 | 56 |
| 81 | Electronic Components and Accessories | 72 | 57 |
| 82 | Miscellaneous Electric Machinery, Equipment & Supplies | 73 | 58 |
| 83 | Motor Vehicles and Equipment | 74 | 59 |
| 84 | Aircraft and Parts | 75 | 60 |
| 85 | Other Transportation Equipment | 76 | 61 |

Code BNL 110 - 110 sector version
BNL 101 - 101 sector version
BEA - Bureau of Economic Analysis, U.S. Dept. of Commerce
Source: (3,p.3)

TABLE 6.2
(Continued)

SECTORAL CLASSIFICATION OF ENERGY INPUT-OUTPUT MODEL
AND ALIGNMENT WITH OTHER CLASSIFICATION SYSTEMS

| BNL 110 | Sector | BNL 101 | BEA |
|---|---|---|---|
| 86 | Professional, Scientific & Controlling Instruments & Supplies | 77 | 62 |
| 87 | Optical, Opthalmic, & Photographic Equipment and Supplies | 78 | 63 |
| 88 | Miscellaneous Manufacturing | 79 | 64 |
| 89 | Railroads and Related Services | 80 | 65.01 |
| 90 | Local, Suburban & Interurban Highway Passenger Transportation | 81 | 65.02 |
| 91 | Motor Freight Transportation and Warehousing | 82 | 65.03 |
| 92 | Water Transportation | 83 | 65.04 |
| 93 | Air Transportation | 84 | 65.05 |
| 94 | Pipeline Transportation | 85 | 65.06 |
| 95 | Transportation Services | 86 | 65.07 |
| 96 | Communications Except Radio & Television Broadcasting | 86 | 67 |
| 97 | Radio and TV Broadcasting | 88 | 67 |
| 98 | Water and Sanitary Services | 89 | 68.01 |
| 100 | Finance and Insurance | 91 | 70 |
| 101 | Real Estate and Rental | 92 | 71 |
| 102 | Hotels & Lodging: Personnel and Repair Service, Except Auto Repair | 93 | 72 |
| 103 | Business Services | 94 | 73 |
| 104 | Automobile Repair and Service | 95 | 75 |
| 105 | Amusements | 96 | 76 |
| 106 | Medical, Educational Services & Nonprofit Inst. | 97 | 77 |
| 107 | Federal Government Enterprises | 98 | 78 |
| 108 | State and Local Government Enterprises | 99 | 79 |
| 109 | Business Travel, Entertainment and Gifts | 100 | 81 |
| 110 | Office Supplies | 101 | 82 |

Code BNL 110 - 110 sector version
BNL 101 - 101 sector version
BEA - Bureau of Economic Analysis, U.S. Dept. of Commerce

Source: (3,p.3)



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